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SO2 REDUCTION IN NON-UTILITY COMBUSTION SOURCES

TECHNICAL AND ECONOMIC COMPARISON OF ALTERNATIVES

BATTELLE COLUMBUS LABORATORIES

PREPARED FOR
INDUSTRIAL ENVIRONMENTAL RESEARCH LABORATORY

OCTOBER 1975

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The report gives results of an analysis of non-utility combustion (NUC) sources for various size classes and fuel types with respect to the significance of SO2 emissions. Technical and economic comparisons of various SO2 control alternatives were made for the important size classes and fuel types. Categories of alternatives included are: physical cleaning of coal, coal gasification, coal liquefaction, fluidized-bed combustion of coal, and flue gas desulfurization. For small size classes of NUC sources, applicabilities of package sorption systems were reviewed.

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IN NON-UTILITY COMBUSTION SOURCES
TECHNICAL AND ECONOMIC COMPARISON OF ALTERNATIVES

by

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ABSTRACT

An analysis of non-utility combustion (NUC) sources was conducted for various size classes and fuel types with respect to the significance of sulfur dioxide emissions. Technical and economic comparison of various sulfur dioxide control alternatives was made for the important size classes and fuel types. Categories of alternatives included in the study are: physical cleaning of coal, coal gasification, coal liquefaction, fluidized-bed combustion of coal, and flue gas desulfurization. For small size classes of NUC sources, applicabilities of package sorption system were reviewed.

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EPA policy is to express all measurements in metric units. When implementing this practice will result in undue cost or lack of clarity, conversion factors are provided for the nonmetric units used in the report. Generally, this report used British unit of measure. For conversion to the metric system, use the following conversion factors.

TABLE OF CONVERSION FACTORS

Multiply English Unit	by Conversion	To Obtain Metric Unit
acres	0.405	hectares
acre-feet	1233.5	cubic meters
barrel, oil	158.97	liters
British Thermal Unit	0.252	kilogram-calories
British Thermal Unit/pound	0.555	kilogram calories/kilogram
cubic feet/minute	0.028	cubic meters/minute
cubic feet/second	1.7	cubic meters/minute
cubic feet	0.028	cubic meters
cubic feet	28.32	liters
cubic inches	16.39	cubic centimeters
degree Fahrenheit	0.555(°F-32)(a)	degree Centigrade
feet	0.3048	meters
gallon	3.785	liters
gallon/minute	0.0631	liters/second
horsepower	0.7457	kilowatts
inches	2.54	centimeters
inches of mercury	0.03342	atmospheres
pounds	0.454	kilograms
million gallons/day	3785	cubic meters/day
mile	1.609	kilometer
pound/square inch (gauge)	(0.06805 psig+1)(a)	atmospheres (absolute)
square feet	0.0929	square meters
square inches	6.452	square centimeters
tons (short)	0.907	metric tons (1000 kilograms)
yard	0.9144	meters

(a) Actual conversion, not a multiplier.

FINAL REPORT

on

SO_2 REDUCTION IN NON-UTILITY COMBUSTION SOURCES--TECHNICAL AND ECONOMIC COMPARISON OF ALTERNATIVES

to

ENVIRONMENTAL PROTECTION AGENCY

by

P.S.K. Choi, E. L. Kropp, W. E. Ballantyne,
M. Y. Anastas, A. A. Putnam, D. W. Hissong,
and T. J. Thomas

from

BATTELLE
Columbus Laboratories

October 1, 1975

INTRODUCTION

The objective of the tasks (EPA Contract No. 68-02-1323, Tasks 13 and 19) was to analyze available small industrial and commercial boiler data and to evaluate various alternatives for the reduction of SO_2 from the non-utility combustion (NUC) sources. This study covered the review of the existing boiler data obtained from various sources such as the National Emissions Data System (NEDS), the Walden Survey, current EPA-related programs, and American Boiler Manufacturers Association data. A methodology was developed for estimation of missing data and various boiler subgroups were evaluated with respect to the significance of SO_2 emissions. The control alternatives under consideration included clean fuels, processed fuels, combustion modification, and flue gas desulfurization (FGD). Additional emphasis was placed on a technologically promising segment of flue gas desulfurization; the segment being package sorption.

The report is thus broken into three parts: Part I describes the acquisition and analysis of boiler data, Part II discusses in general the four control alternatives studied, and Part III provides in-depth background information to assist EPA in preparing development/demonstration studies on package sorption systems for the abatement of SO_x emissions from small industrial and commercial boilers.

MANAGEMENT SUMMARY

This study was performed for the Environmental Protection Agency (EPA) under Contract No. 68-02-1323, Tasks 13 and 19. The objective of Task 13 was to analyze available small industrial and commercial boiler data and to evaluate various alternatives for the reduction of sulfur dioxide emissions from the non-utility combustion (NUC) sources. The objective of Task 19 was to develop background information which could assist EPA in preparing development/demonstration studies on package sorption systems for the abatement of sulfur dioxide from the NUC sources under consideration.

This report consists of three parts. The acquisition and analysis of the boiler data are described in Part I. Data on existing industrial and commercial boilers were obtained, for the most part, from the National Emissions Data System (NEDS), the Walden survey, current EPA-related programs, and American Boiler Manufacturers Association (ABMA) compilations. Where gaps in the data base existed, methodology was developed for estimation of key data. Various boiler classes were also evaluated with respect to the significance of sulfur dioxide emissions.

Various control alternatives were reviewed, analyzed, and evaluated and the results are included in Part II. The alternatives under consideration included clean fuels, processed fuels, combustion modification, and flue gas desulfurization (FGD). A set of evaluation criteria was established and employed to examine the technical feasibility of each alternative. The control costs were estimated for the boiler classes of environmental concern in order to examine the economic feasibility of the alternatives. In the evaluation study, attempts were made to provide the general technical and economic information for the selected boiler classes using the average parameter values of the acquired boiler data. Specific case studies were not made.

Part III of this report covers the survey and evaluation of existing and potential package sorption devices and technologies applicable to the NUC sources identified in Part I. The sorption processes considered in this study included both liquid and solid phase sorption processes which were classified into two categories--throwaway and regenerable processes.

The results of the boiler data analysis indicated that the parameters that fix the annual load factor are the use category and type of fuel fired. The annual load factor is higher for coal and industrial use than for oil and commercial use. The majority of the flue gas temperatures are between 400° and 450° F. There is a tendency for the temperature to be higher for high-sulfur oil-firing units and for coal-firing units of smaller size. The important boiler size classes to consider for reduction of area source emissions are middle to large size coal units (both stoker and pulverized) and almost all classes burning high-sulfur oil (including sizes as small as 2×10^6 Btu/hr).

It is expected that the NUC sources under consideration will be forced to use dirty fuels (i.e., high-sulfur fuels) due to the insufficient supply of clean fuels. Processed fuels such as high-Btu synthetic natural gas (SNG) and coal liquefaction products produced on a large scale are economically favorable over application of FGD processes for the small size classes of the NUC sources. The primary concern for the processed fuels, however, is the availability of the processes in the near future.

Technically, all FGD processes are feasible for application to the NUC sources under consideration, although the high oxygen concentration in the flue gas may possibly cause difficulties in processes where oxidation is undesirable. The FGD processes are economically favorable over other alternatives for the NUC sources in large size classes. Most important of all, many FGD processes are commercially available although the application to the NUC sources remains to be demonstrated.

Among various FGD processes, regenerable processes with regeneration performed at a central facility, such as the MgO process with central regeneration, should be the most attractive choice where there is no existing capability for regeneration within the NUC source physical plant. Such a system is relatively simple in operation, small in size, and low in capital and annualized costs. The sorption capacity of the sorbent, however, should be high so that the size of the sorption unit and the quantity of sorbent to be used may be minimal.

Throwaway processes should be more favored than integrated regenerable processes in controlling NUC sources without an in-house capability for regeneration if there is land available for the sludge disposal. This is mainly because the former generally is simpler in operation, smaller in size, and lower in annualized cost than the latter.

Currently no package sorption system is available for the control of SO_2. The package unit concept for various FGD processes is not deemed feasible for control of small and large sizes of NUC sources. This is mainly because for small size NUC sources, the economics are very favorable toward processed fuels, and, thus, the FGD processes are not economically feasible as compared with processed fuels. For large size NUC sources, the FGD processes should become economically favored over other alternatives; however, the size of the system would become too big to be handled as a package unit. The concept of a packaged unit may be feasible for control of medium size NUC sources, i.e., boiler sizes up to 30,000 lb/hr, if the packaged unit can be manufactured for a low cost and installed relatively easily, and the fixed capital charge portion of the annualized cost can be kept to a small fraction of the capital cost.

CONCLUSIONS

(1) The annual load factor was computed for each boiler for which data were available by dividing the total fuel heat input rate per year by the design firing rate and the number of hours per year. A consideration of the load factors led to the conclusion that, for the range of design capacities from 10^6 Btu/hr through 5×10^8 Btu/hr, the parameters that fix the annual load factor are the use category and type of fuel fired (with sulfur content not important).

(2) NEDS data on flue gas flow rate for individual boilers can be justified; however, it is not possible to make a definitive statement about any average flue gas flow rates due to the many different types of combustion equipment using different amounts of excess air.

(3) An assessment of reported stack exit temperatures leads to the conclusion that a majority of temperatures are between 400°F and 450°F. There is a tendency for the temperature to be higher for high-sulfur residual-oil-firing units and for coal-fired units of smaller size.

(4) Typical sulfur contents in fuel ranged from nil in natural gas to a high of almost 3.5 percent in stoker-fired coal units. Coal-fired units showed a slight increase in fuel sulfur content as boiler size increased, but all other fuels fired tended to remain at a constant sulfur level.

(5) Sulfur oxide emissions were calculated using NEDS analysis and EPA emission factors. Both potential and actual (adjusted by annual load factor) emissions were calculated. Conclusions reached are that the important boiler size classes to consider for reduction of non-utility combustion source emissions are middle to large size (10×10^6 to 500×10^6 Btu/hr) coal units (both stoker and pulverized) and almost all classes burning high-sulfur residual oil (including sizes as small as 2×10^6 Btu/hr).

(6) The ABMA data file can be used successfully to supplement the NEDS file in the analysis of boiler data.

(7) The use of clean fuels, such as natural gas, distillate fuel oil, low-sulfur residual oil, and low-sulfur coal are the best control

alternatives for the non-utility combustion sources under consideration if the clean fuels are available and can be obtained for a reasonable price. The present uncertainties in the energy area, however, make projections of fuel supply and cost very difficult and subject to considerable speculation. Particularly important are the questions of whether the price of natural gas will be deregulated and to what extent the U. S. will continue to import foreign oils.

(8) Processed fuels produced on a large scale are economically favored over application of FGD processes to boilers in the small size ranges of the non-utility combustion sources. The capital cost requirement to the boiler system is low for the processed fuels. The annualized cost is also low, mainly due to the small fixed capital charges. The primary concern associated with processed fuels, however, is the availability of the processes in the near future.

(9) Technically, all FGD processes are feasible for application to commercial and industrial boilers, although the high oxygen concentration in the flue gas may possibly cause difficulties in processes where oxidation is undesirable such as the double alkali and Wellman-Lord processes. The FGD processes are economically favorable over other alternatives for the non-utility combustion sources in large size classes. Most important here, many FGD processes are commercially available, although the application to the non-utility combustion sources remains to be demonstrated.

(10) Physical cleaning of coal is very attractive in its economics; however, its application is limited to certain types of coal.

(11) A coal gasification process for low-Btu gas generation should be a retrofit system to a group of non-utility combustion sources. The application in general is limited to coal-fired boilers due to non-compatible combustion chamber configurations of gas- and oil-fired boilers.

(12) A coal gasification process for high-Btu synthetic natural gas manufacture will be a viable alternative for the non-utility combustion sources except for the large size boilers. The process will not be available until 1980 to 1983.

(13) The solvent refined coal process is a viable alternative for control of non-utility combustion sources. The solid product will be

used for coal-fired boilers and the liquid product for oil-fired boilers. For oil-fired boilers, however, the H-coal process will be more favorable economically since heating of the product from the H-coal process in its utilization is required to a lesser extent. The liquefaction process, however, will not be available until 1981 to 1983.

(14) Among various FGD processes, regenerable processes with regeneration performed at a central facility, such as the MgO process with central regeneration, should be the most attractive choice where there is no existing capability for regeneration within the non-utility combustion source physical plant. Such a system is relatively simple in operation, small in size, and low in capital and annualized costs. The sorption capacity of the sorbent, however, should be high, so that the size of the sorption unit and the quantity of sorbent to be used may be minimal.

(15) Throwaway processes could be more favorable than integrated regenerable processes in controlling non-utility combustion sources where there is no in-house capability for regeneration and no close central regeneration capability. This is mainly because the former generally is simpler in operation, smaller in size, and lower in annualized cost than the latter. In addition, the operation of a throwaway process becomes more reliable as more information is developed for the underlying chemistry of the process. Although land is required for sludge disposal, the solid waste generated by the processes is small in quantity because of the small system size.

(16) Integrated regenerable processes may be more favorable in cases of industrial boilers where there are captive uses for sulfur compounds for in-house capability for regeneration. For example, the pulp and paper industry can regenerate spent sodium sulfite or ammonium scrubbing solution in its manufacturing operation. The chemical industry frequently has captive uses for sulfuric acid or sulfate salts. The petroleum industry has sources of hydrogen sulfide that could be used to produce sulfur from SO_2 emissions. An anlysis would be needed for the specific individual case to determine the technical and economical feasibilities of the various FGD processes.

(17) Currently no package sorption system is available for the control of SO_2. The package unit concept for various FGD processes is not deemed feasible for control of small and large sizes of non-utility combustion sources. This is mainly because for small size non-utility combustion sources, the economics are very favorable towards processed fuels, and, thus, the FGD processes are not economically feasible as compared with processed fuels. For large size non-utility combustion sources, the FGD processes become economically favorable over other alternatives; however, the size of the system would become too big to be handled as a package unit. The concept of a packaged unit may be feasible for control of medium size non-utility combustion sources, i.e., boiler sizes up to 30,000 lb/hr, if the packaged unit can be manufactured for a low cost and installed relatively easily, and if the fixed-capital portion of the annualized cost can be kept to be a small fraction of the capital cost.

RECOMMENDATIONS

(1) The existing NEDS-ABMA data files should be kept current to maintain trend analysis capabilities.

(2) Non-utility combustion source SO_x emissions appear to be concentrated in coal-fired units of between 2×10^7 and 5×10^8 Btu/hr size class and in high sulfur residual oil-fired units of sizes between 2×10^6 and 5×10^8 Btu/hr; thus, SO_x control technique analyses should be performed for these size range units.

(3) Clean fuel development activities should be accelerated to promote control of small non-utility combustion sources more economically in the near future. Priorities should be given to high-Btu coal gasification, solvent refined coal, and H-coal processes. Development of physical and chemical coal cleaning processes should be continued so that the sulfur removal efficiency may be improved to a high degree (i.e., greater than 70 percent of total sulfur) and the application of these processes can be broadened to many types of high-sulfur coal.

(4) Conceptual and demonstration studies of FGD processes for application to the non-utility combustion sources are recommended under the following categories:

(a) Regenerable processes with central regeneration facility

(b) Throwaway processes.

The selection of the regenerable process will be based on the magnitude of sorbent capacity, ease of transportation of spent and regenerated sorbent material, and simplicity of sorption operation. Magnesium oxide and sodium sulfite-based regenerable processes will be appropriate due to the high sorption capacity. The selection of the throwaway process will be based on availability of raw material, reliability of operation, and simplicity of the system. Limestone based simple wet scrubbing and double alkali processes are recommended.

(5) A study is recommended to identify sorbents which have a high sorption capacity for SO_2. A literature survey will be necessary to

review various sorbent materials. The result of this study will be useful to select sorbents for regenerable processes with central regeneration facilities.

(6) No study is recommended in the immediate future for the concept of packaged sorption units for non-utility combustion sources. The study should be carried out when the results of the conceptual and demonstration studies of the selected FGD processes mentioned above are available.

PART I

INDUSTRIAL AND COMMERCIAL BOILER DATA BASE

DESCRIPTION OF DATA SOURCES

The primary sources utilized in acquiring industrial and commercial boiler data were the National Emission Data System (NEDS) and American Boiler Manufacturers Association (ABMA) data. Data were accumulated on all sizes of commercial, industrial, and utility boilers; utilities were included since utility boiler information was contained in the data files analyzed.

The NEDS file contains pollutant source information gathered primarily in 1972 and 1973. The information on file includes the following parameters:

- State
- County
- ACQR
- Plant ID Number
  - City
  - UTM Zone
- Year of Record
- Establishment Name and Address
- Person to Contact
- Owner
- Point ID
- Standard Industrial Classification (SIC)
- IPP Process
- UTM Coordinates
- Stack Data
- Boiler Design Capacity
- Control Equipment
- Estimated Control Efficiency, percent
- Percent Annual Throughput
- Normal Operating Time, Hours
- Emission Estimate, tons/yr
- Percent Space Heat
- Allowable Emissions, tons/yr
- Compliance Status
- Control Regulations
- Source Classification Code (SCC)
- Fuel, Process, Solid Waste Operating Rate
- Maximum Design Rate
- Sulfur Content, percent
- Ash Content, percent
- Heat Content, 10^6 Btu/scc, i.e., 10^6 Btu/10^6 scf for gas, 10^6 Btu/10^3 gal for oil, and 10^6 Btu/ton for coal

These data are on file for all sources inventoried from 1972 to mid-1974. The year for which the data apply is recorded and varies from 1968 through 1974.

The American Boiler Manufacturers Association (ABMA) records data on all boilers sold by its members. The data on watertube boilers are kept in computer-card form, one card prepared for each boiler sold. Records are also kept for firetube (Scotch) boilers but the records are not nearly as detailed as are the watertube records.

The watertube computer file was initiated in 1965 and is currently updated each month. Information contained in the file is as follows:

Capacity per Unit:
  Boiler capacity reported in thousands pounds of steam per hour* as the maximum capacity on the base fuel.

Primary Fuel:
  Bituminous Coal
  Oil
  Natural Gas
  Wood Bark, or Solid Wood
  Bagasse
  Black Liquor
  Other Fuels
  Waste Heat
  Waste Heat, Auxiliary firing
  Lignite
  Raw Municipal, Unsorted
  Raw Municipal, Non-combustible Removed
  Raw Municipal, Sorted & Sized
  Other Industrial Waste

Alternate or Auxiliary Fuel

Firing Method:
  Pulverized Coal
  Spreader Stoker
  Underfeed Stoker
  Overfeed Stoker
  Other Fuel Firing
  Gas Turbine or Engine Exhaust
  Other Non-combustible Waste Gas
  Combustible Waste Gas
  Non-solid Fuel Firing

Packaged or Field Assembled:
  Pressure vessel completely shop assembled.
  Pressure vessel shop assembled and shipped as two, three, four, five, or six major modules.
  Packaged design shipped knocked-down.
  Field assembled, bottom supported.
  Field assembled, top supported

Standard Industrial Classification Number, 2 digit.

Domestic or Export.

Stationary or Marine Boiler.

---

* 1000 Btu/hr = 1 lb steam/hr.

Type Code:
  Watertube
  Wasteheat, watertube, bare tube
  Wasteheat, watertube, extended surface
Draft Conditions:
  Pressurized furnace
  Balanced draft
Steam or Hot Water
Capacity per Unit (MW):
  For utility boilers, the manufacturers rating in MW of the generator is to be used with the boiler as indicated. If the boiler is a non-generating unit, the designation "9999" is indicated.
Design Pressure, psig
Operating Pressure, psig
Saturated Steam or Hot Water Outlet Temperature
  Saturated Steam
  Hot Water Outlet Temperature, °F
Superheat, °F
Steam Temperature at First Reheater Outlet, °F
Steam Temperature at Second Reheater Outlet, °F.

The aforementioned data are recorded for each watertube boiler sold; however, there are no data recorded regarding a projected installation date or specific location for the boiler.

Firetube (Scotch) boiler data are recorded mainly in the form of number of boilers sold each month. Number sold is recorded for low-pressure steam, high-pressure steam and hot-water boilers. Fuel (gas, oil, or combination gas and oil) is recorded but not for a specific boiler, e.g., of the total number of LP, HP, and HW boilers sold, it is possible to determine that a given percentage of all boilers burned oil but not that a given percentage of only LP boilers burned oil. None of the firetube data are currently computerized.

The two main differences between the files are (1) the ABMA data concerns boilers sold while the NEDS data concerns boilers in the field, and (2) The ABMA file is a boiler information file whereas NEDS is a "source" registration file which contains information on emission sources but not primarily on boilers. Because boiler information is not stressed heavily in NEDS, the accuracy of the boiler information is unknown. The data are probably good for both large boilers and large companies since large companies tend to supply more knowledgeable people to fill out emission inventory questionnaires and data on large boilers are usually more fully documented. However, as the boiler sizes get smaller and the company size gets smaller, the information file probably tends to become more and more suspect. Other sources of boiler data, such as the Walden Survey(1) and various trade journals were reviewed in addition to using the two computer files to characterize the boiler population.

DESCRIPTION OF PARAMETERS NECESSARY TO CHARACTERIZE BOILER POPULATION

In order to estimate SO_2 control costs or even decide on a feasible control technology, one must first be able to quantify the pertinent characteristics of the SO_2 source. Several boiler parameters become important in this quantification process, thus the need to define classes of boilers by these parameters.

In general, one would like to maintain as few classes of parameters as possible in any analysis to reduce the amount of material that must be comprehended, and to increase the count number in each class and, thus, the accuracy of any deductions about the parameter. On the other hand, any factor which might cause a decided change in pertinent characteristics of a class must be used to split the class. For the present study, as a result of both the viewpoint just presented and the results presented in Reference (2), the following parameters have been considered.

Use Category

Three use categories were considered: commercial, industrial, and utility. While it has been common to use arbitrary design capacity for a dividing line between use categories, this study has shown that the annual load factor is more dependent on use category than on design size, or even fuel. If one were to calculate a load factor for several given size classes of boilers, irrespective of use category, one would see an apparent increase in load factor with increasing design size. The NEDS data, however, when subdivided by use categories, shows that the load factor is independent of design size but dependent on use category. The commercial boilers have a lower load factor over almost all design size ranges than do industrial boilers, which in turn have a lower load factor than utility boilers. An average load factor over an active boiler class will appear to increase with size simply due to the mix (number) of boilers in the classes.

Size

In Reference (2), in classifying data as a function of boiler size, the total population of boilers was divided into 5 size ranges (in 10^6 Btu/hr steam) as follows: 10^+ - 20, 20^+ - 50, 50^+ - 100, 100^+ - 200, 200^+ - 500. To maintain consistency in form, the lower limit in each range, including the lowest range, starts slightly above the division, e.g., 10,001, 20,001, etc. Divisions at these points were found to be consistent with the literature. In the present study, this range is extended to cover the entire range of NEDS data. The assumption is made that ABMA data on lb/hr of steam can be converted to Btu/hr by use of a factor of 1000.

Table 1 shows the total classification used to cover the NEDS data. In addition to the nominal ranges, as listed, the geometric mean size is presented. Class 1* is a result of the NEDS reporting format, i.e., the smallest size able to be recorded is 1×10^6 Btu/hr and only boilers in the size range $.5 \times 10^6$ - 1.4×10^6 Btu/hr may be included in the NEDS as a 1×10^6 Btu/hr boiler. It was possible to identify numbers of boilers and total installed capacity for each fuel type in Class 1* but the fuel properties, i.e., sulfur and ash content, etc., were assumed to be the same as for Class 1.

TABLE 1. SIZE RANGES USED TO CLASSIFY NEDS DATA

Size Class Number	Nominal Range of Capacity, 10^6 Btu/hr	Log Mean Size,(c) 10^6 Btu/hr
1*(a)	1	.837
1(b)	$1^+ - 2$	1.936
2	$2^+ - 5$	3.674
3	$5^+ - 10$	7.563
4	$10^+ - 20 \times 10$	14.640
5	$20^+ - 50 \times 10$	32.140
6	$50^+ - 100 \times 10$	71.210
7	$100^+ - 200 \times 10^2$	141.900
8	$200^+ - 500 \times 10^2$	316.700
9	$500^+ - 1000 \times 10^2$	707.600
10	$1000^+ - 2000 \times 10^3$	1415.000
11	$2000^+ - 5000 \times 10^3$	3163.000
12	$5000^+ - 10,000 \times 10^3$	7072.000
13	$10,000^+ - 20,000 \times 10^4$	14140.000

(a) 1* does not refer to size range, but to a specific set of data with minimum recordable capacity.

(b) Because of the incremental values of 1, 2, 3, this set of data contains only data of labeled capacity 2 x 10^6 Btu/hr.

(c) Square root of the product of maximum size in range (to nearest .1 x 10^6 Btu/hr) and minimum size in range (to nearest .1 x 10^6 Btu/hr).

Fuel Type

The fuel divisions of main interest to this project are coal, residual oil, distillate oil, and gas. Because of the differences in firing methods and thus in the amount of excess air, however, the coal types are split into pulverized fuel and cyclone in one classification, and all other systems (stoker types) in the second classification. An examination of the data has also shown that non-natural-gas data appear to be highly scattered. Therefore, the gas data are divided into two classes, natural gas (including LPG) and non-natural gas. This results in six categories as shown in Table 2.

TABLE 2. COMPILATION OF FUEL TYPES

Pulverized coal and cyclone fired coal
Stoker and other coal not in item above
Residual oil
Distilled oil
Natural gas and LPG
Other gases not in item above

Annual Load Factor

The annual load factor was computed for each boiler for which NEDS data were available by dividing the total fuel heat input rate per year by the design firing rate and the number of hours per year. In several instances, this value was greater than unity, often by a considerable amount. This indicated some error in these data, so the value of 1.05 was arbitrarily assigned. This will result in some overestimation, but neglect of these units will result in too low a value since some boilers are fired at above the design rate. Comparison of the portion of the data developed for this study with the data of Reference (4) indicates only about 5 percent difference in annual load factor for each class. For each category defined by size class, use category, fuel, and sulfur content, an average annual load factor was calculated.

A consideration of the annual load factor values for each category, along with the number of boilers and total capacity involved in defining each value, led to the conclusion that, in most instances, the annual load factor was not a function of design firing rate of the boiler, within the accuracy to which it could be determined. Figure 1 and Table 3 present some of the results of the individual class calculations and show that the load factor is independent of design size class for most cases. In cases of extremely high or low load factors, there are few boilers in the class from which to calculate an average so that any anomalies in the data will have a greater than normal effect on the average.

Average load factor values were calculated for all fuel types and size classes based on the NEDS. These averages are shown in Table 4. The same average load factor results were obtained by either number weighting or total design-firing-rate weighting the annual load factor for each category. In the interest of accurate cost calculations for Parts II and III of this report, and because they were already calculated, it was decided to use specific load factor values from Table 3 instead of the average values from Table 4 for the cost equations. For future work, however, the conclusion formed was that the values from Table 4 will be sufficient for costing purposes.

In summary, it was found that for most practical purposes, for the range of design capacities considered herein, from 10^6 Btu/hr through 2×10^{10} Btu/hr, the parameters that fix the annual load factor are the category of use, and the type of fuel (with sulfur content not important).

Stack Temperature

The stack temperatures shown in Table 5 and Figure 2 are the average temperatures from NEDS for size classes 1 through 8 (non-utility combustion source classes). They are defined as "the temperature of the exhaust stream at the stack exit, in degrees Fahrenheit, under normal operating conditions. If measured temperatures are not available, an estimate to the nearest 50°F should be made."

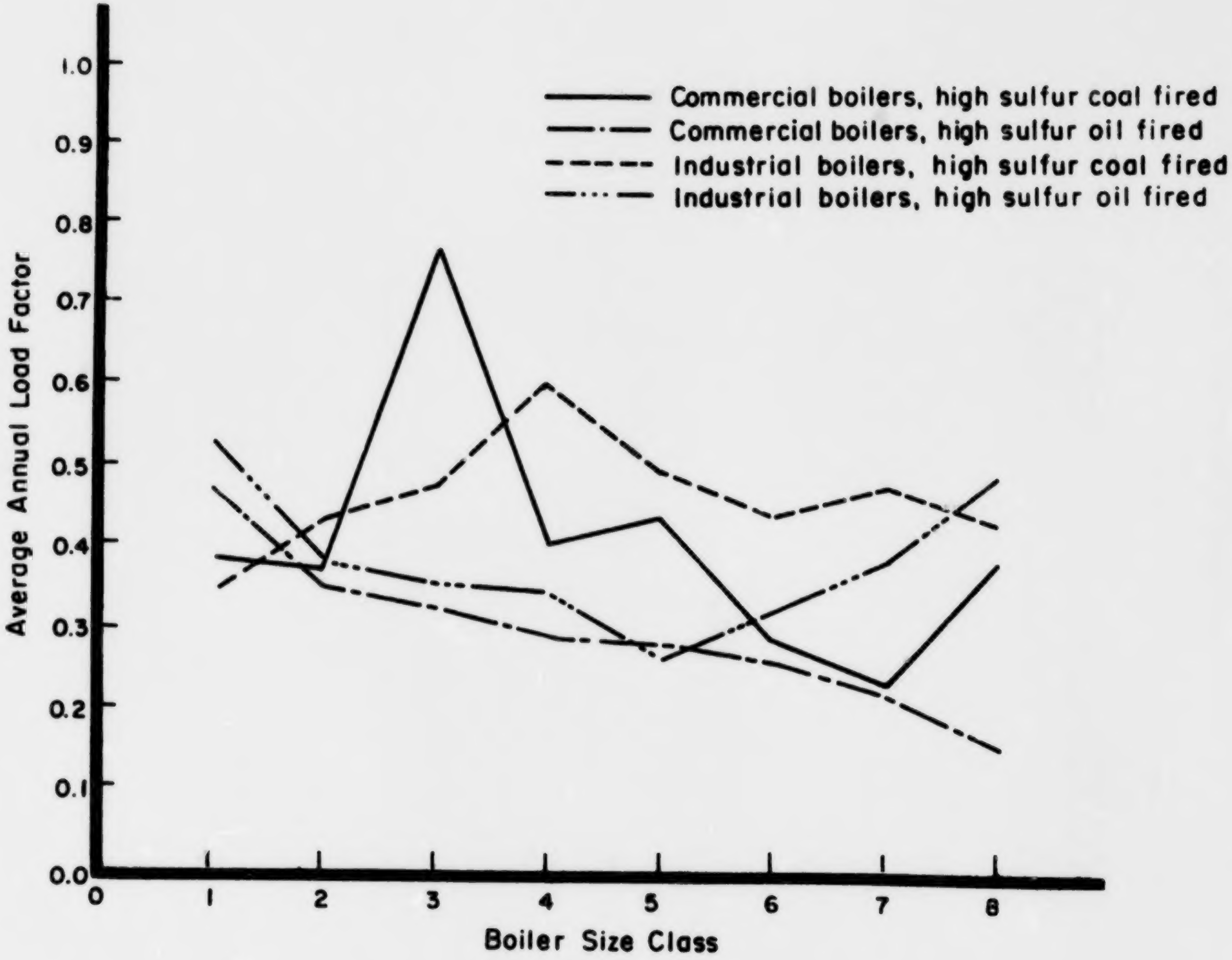

FIGURE 1. AVERAGE ANNUAL LOAD FACTOR FOR HIGH SULFUR COAL FIRED AND HIGH SULFUR OIL FIRED BOILERS IN EIGHT NON-UTILITY COMBUSTION SOURCE SIZE CATEGORIES

TABLE 3. ANNUAL AVERAGE LOAD FACTOR FOR HIGH SULFUR COAL AND OIL-FIRED BOILERS IN 8 AREA SOURCE SIZE CLASSES

Size Class	Fuel Type	Load Factor	
		Commercial	Industrial
1	High Sulfur Coal	.381	.339
2		.371	.415
3		.752	.475
4		.393	.599
5		.417	.497
6		.292	.417
7		.228	.428
8		.381	.411
1	High Sulfur Oil	.459	.523
2		.350	.372
3		.302	.357
4		.279	.351
5		.276	.263
6		.259	.338
7		.223	.383
8		.149	.430

TABLE 4. ANNUAL LOAD FACTORS AS A FUNCTION OF USE AND FUEL TYPE, BASED ON NEDS DATA

	Commercial(e)	Industrial	Utility
Stoker coal(a)	0.305	0.426(g)	0.479
Pulverized coal(b)	0.424	0.524(g)	0.423
Residual oil	0.245	0.368	$0.429_{LS}/0.647_{HS}$(i)
Distillate oil	0.206	0.330(h)	0.130(j)
Natural gas(c)	0.318	0.518	0.474
Other than natural gas(d)	--	0.630	(k)

(a) Stoker coal includes all coal firing except pulverized coal and cyclone.

(b) Pulverized coal includes cyclone.

(c) Natural gas includes natural gas and LPG. All low sulfur.

(d) Other than natural gas includes all gases except natural gas and LPG. Almost all low sulfur.

(e) Load factor values valid for sizes in range $1^{+} \times 10^{6}$ to 5×10^{8} Btu/hr. Values ranged from 0.01 to 0.16 for the few boiler data above this design size range.

(f) Up through 5×10^{8} Btu/hr. For the few boilers above, use 0.268.

(g) Up through 5×10^{8} Btu/hr. For boilers above, use 0.580.

(h) No data above 5×10^{8} Btu/hr.

(i) Use industrial value up to 5×10^{8} Btu/hr.

(j) All low sulfur.

(k) Very sparse population; suggest using industrial value.

TABLE 5. AVERAGE TEMPERATURE OF STACK GASES IN °F AS A FUNCTION OF FUEL AND NON-UTILITY COMBUSTION SOURCE SIZE CATEGORY

	Size Class[1]							
Fuel	1	2	3	4	5	6	7	8
Coal	535	489	460	439	470	438	406	410
	403	482	423	464	495	450	410	392
Residual	338	370	360	392	416	427	418	421
	555	413	431	452	454	453	430	406
Distillate	411	501	418	390	429	444	491	378
	354	387	343	320	503	346	463	--
Gas	377	420	442	430	457	460	405	435
	--	--	500	350	721	560	545	850

(1) Upper number is for low sulfur, and lower number is for high sulfur.

An observation of the data, together with a consideration of the number of items involved in each point, indicates that the majority of temperatures are between 400°F and 450°F. There is a tendency for the temperature to be higher for the high-sulfur residual than for the low-sulfur residual; this is probably the result of a deliberate effort to avoid acid condensation in the stack. The high values for high-sulfur gas are associated with a small number of special gases, not natural gas. For coal-fired units of smaller sizes (i.e., Classes 1-5), the temperature tends to fall in the 450-500°F range rather than 400-450°F range.

Fuel Sulfur Content

Following categorization of a boiler with respect to size and fuel type, the influence of fuel sulfur content (weight percent) was examined. Two categories were selected, less than or equal to 1.0 percent sulfur (defined as low sulfur) and greater than 1.0 percent sulfur (defined

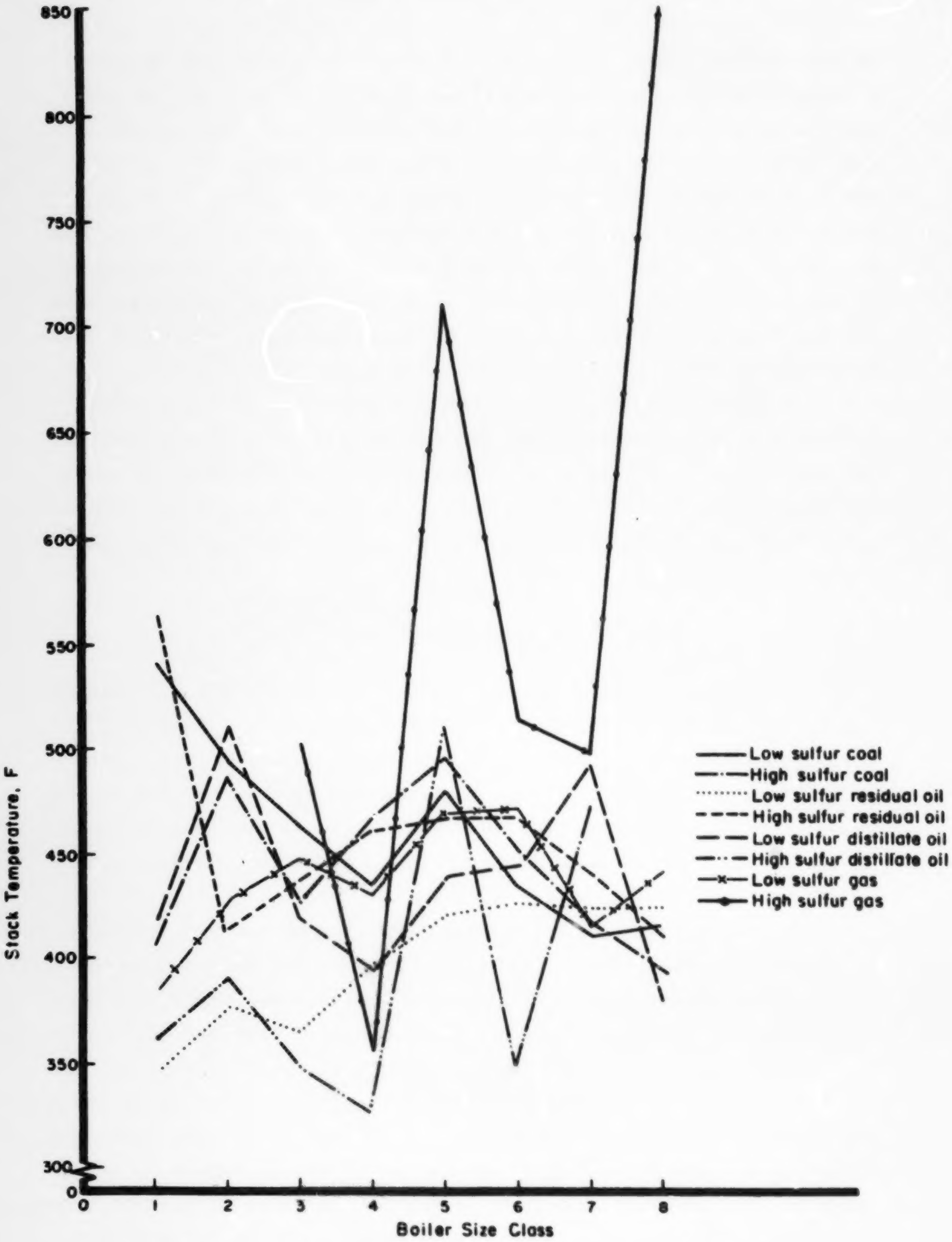

FIGURE 2. AVERAGE TEMPERATURE OF STACK GASES (^{o}F) AS A FUNCTION OF FUEL AND NON-UTILITY COMBUSTION SOURCE SIZE CATEGORY

as high sulfur). Since the sulfur content was to be used for the purpose of calculating control costs on boilers, it was decided to use the modal (most frequently occurring value) sulfur content rather than an average. This would assure that costs would be calculated for fuel with sulfur content more nearly like that of the fuels received by a boiler owner. The mode was thus calculated for each category defined by size, fuel type, and low or high sulfur group. There appears to be little relationship between sulfur content modes in each class, i.e., sulfur content seems to remain independent of use category. The only discernible trend is a slight tendency of sulfur content to increase with boiler size increases.

Sulfur content, in this study, is a means to an end (control costs); therefore, it was decided that further effort on the analysis of the parameter, i.e., graphing it, would not be fruitful. Table 6 presents the modal fuel sulfur content for each of 13 boiler size classes, six fuel types (with low and high sulfur considered separately), and three use categories.

Flue Gas Flow Rate

Initially, an attempt was made to analyze the flue-gas flow rates, as reported in NEDS, with respect to design firing rate. Unexpectedly, analysis of the distribution of flow rates with respect to design firing rate indicated that there was little effect of design firing rate on flue gas rates. Therefore, the flow rate data were split into classes defined by boiler fuel type and use category and then normalized with respect to design firing rate.

The object of all of the data manipulation was to determine if one could define a "typical" flue gas flow rate given boiler parameters such as size, use category, and fuel type. Further analysis of the data following normalizing and different class divisions indicated that one could not define a "typical" flue gas flow rate due to the wide variation in reported flow rates for a given class of boilers. It was decided that an investigation was necessary to determine the reason for the wide variation in the reported NEDS flue gas flow rates. Upon examination of several references in conjunction with the NEDS data, it was found that at least six factors influence the range of reported flow rates. These factors include (1) stoichiometric

TABLE 6. MODAL FUEL SULFUR CONTENT, WEIGHT PERCENT(1)

Fuel	Class 0: 5^{+}-10 x 10^5 Btu/hr			Class 1: 1^{+}-2 x 10^6 Btu/hr			Class 2: 2^{+}-5 x 10^6 Btu/hr			Class 3: 5^{+}-10 x 10^6 Btu/hr		
	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility
Stoker Coal,												
Low Sulfur	0.64	0.75	0	0.64	0.75	0	0.65	1.00	0	0.65	0.98	0
High Sulfur	2.00	1.01	2.5	2.00	1.01	2.5	3.40	3.28	3.25	3.22	2.50	0
Pulverized Coal,												
Low Sulfur	0.65	0.65	0	0.65	0.65	0	--	0.65	0	0.65	0.85	0
High Sulfur	0	3.50	0	0	3.50	0	0	4.00	0	0	0	0
Residual Oil,												
Low Sulfur	1.00	0.96	0.75	1.00	0.96	0.75	0.99	0.96	0.95	0.98	0.97	0.94
High Sulfur	2.29	2.41	2.5	2.29	2.41	2.5	1.01	2.21	2.5	2.23	2.29	2.5
Distillate Oil,												
Low Sulfur	0.26	0.23	0.1	0.26	0.23	0.1	0.25	0.24	0.1	0.25	0.24	0.1
High Sulfur	1.01	1.01	0	1.01	1.01	0	1.01	1.01	0	1.01	1.01	0
Non-Natural Gas,												
Low Sulfur	0	0	0	0	0	0	0	0	0	0	0	0
High Sulfur	0	0	0	0	0	0	0	0	0	0	2.50	0
Natural Gas,												
Low Sulfur	0	0	0	0	0	0	0	0	0	0	0	0
High Sulfur	0	0	0	0	0	0	0	0	0	0	1.	0

(1) Top number is low sulfur value, bottom number is high sulfur value.

TABLE 6. MODAL FUEL SULFUR CONTENT, WEIGHT PERCENT (Continued)

Fuel	Class 4: 1^+-2 x 10^7 Btu/hr			Class 5: 2^+-5 x 10^7 Btu/hr			Class 6: 5^+-10 x 10^7 Btu/hr			Class 7: 1^+-2 x 10^8 Btu/hr		
	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility
Stoker Coal,												
Low Sulfur	0.64	0.91	0.54	0.65	0.75	0.65	0.96	0.95	0.96	0.45	0.98	0.99
High Sulfur	2.86	2.46	1.01	2.58	2.47	2.43	2.54	2.56	3.14	3.20	2.85	2.0
Pulverized Coal,												
Low Sulfur	--	0.91	0.1	0.83	0.75	0.75	0.83	0.93	0.85	0.75	0.93	0.93
High Sulfur	1.01	3.43	2.5	3.50	2.13	1.01	1.01	2.53	3.45	2.00	2.45	2.39
Residual Oil,												
Low Sulfur	0.99	1.00	0.45	0.99	0.99	0.45	0.99	1.00	0.55	0.93	0.99	0.75
High Sulfur	2.27	2.28	2.47	2.06	2.37	2.46	2.31	2.39	2.41	2.20	2.38	2.44
Distillate Oil,												
Low Sulfur	0.25	0.22	0.2	0.25	0.22	0.15	0.27	0	0.23	0.25	0.15	0.25
High Sulfur	1.01	1.01	0	1.01	2.31	0	1.01	1.01	2.5	1.01	1.01	0
Non-Natural Gas,												
Low Sulfur	0	0	0	0	0	0	0	0	0	0	0	0
High Sulfur	0	0	0	0	1.01	0	0	1.01	0	0	1.01	0
Natural Gas,												
Low Sulfur	0	0	0	0	0	0	0	0	0	0	0	0
High Sulfur	0	3.50	0	0	4.00	0	0	1.01	0	0	1.01	0

TABLE 6. MODAL FUEL SULFUR CONTENT, WEIGHT PERCENT (Continued)

Fuel	Class 8: 2^{+}-5 x 10^{8} Btu/hr			Class 9: 5^{+}-10 x 10^{8} Btu/hr			Class 10: 1^{+}-2 x 10^{9} Btu/hr		
	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility
Stoker Coal,									
Low Sulfur	0.73	0.65	0.1	0	0.35	0.75	0	0.75	0
High Sulfur	3.25	1.01	3.0	0	2.33	3.67	0	3.5	4.0
Pulverized Coal,									
Low Sulfur	--	0.76	0.92	0	0.86	0.92	0	0.85	0.93
High Sulfur	0	2.22	2.96	0	3.29	2.39	0	3.33	2.64
Residual Oil,									
Low Sulfur	0.96	0.44	0.85	0.90	0.97	0.85	0	0.45	0.44
High Sulfur	1.01	2.43	2.34	0	2.40	2.26	0	2.5	1.01
Distillate Oil,									
Low Sulfur	0.25	0.15	0.24	0.20	0.15	0.25	0	0	0.95
High Sulfur	0	0	0	0	0	2.50	0	0	0
Non-Natural Gas,									
Low Sulfur	0	0	0	0	0	0	0	0	0
High Sulfur	0	3.50	1.01	0	1.01	0	0	1.01	0
Natural Gas,									
Low Sulfur	0	0	0	0	0	0	0	0	0
High Sulfur	0	0	0	0	0	0	0	0	0

TABLE 6. MODAL FUEL SULFUR CONTENT, WEIGHT PERCENT (Continued)

Fuel	Class 11: 2^{+}-5 x 10^{9} Btu/hr			Class 12: 5^{+}-10 x 10^{9} Btu/hr			Class 13: 1^{+}-2 x 10^{10} Btu/hr		
	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility
Stoker Coal,									
Low Sulfur	0	0	0	0	0	0	0	0	0
High Sulfur	0	3.5	4.0	0	0		0	0	0
Pulverized Coal,									
Low Sulfur	0	0	0.86	0	0	0.56	0	0	0
High Sulfur	0	1.01	2.69	0	5.5	2.76	0	0	4.5
Residual Oil,									
Low Sulfur	0	0.45	0.99	0	0	0	0	0	0
High Sulfur	0	2.5	2.28	0	0	2.5	0	0	0
Distillate Oil,									
Low Sulfur	0	0	0.35	0	0	0.65	0	0	0.25
High Sulfur	0	0	0	0	0	0	0	0	0
Non-Natural Gas,									
Low Sulfur	0	0	0	0	0	0	0	0	0
High Sulfur	0	0	0	0	0	0	0	0	0
Natural Gas,									
Low Sulfur	0	0	0	0	0	0	0	0	0
High Sulfur	0	0	0	0	0	0	0	0	0

firing rate, (2) excess air, (3) leakage, (4) corrections for stack temperature, (5) the boiler load factor, and (6) the fan flow safety margin. The combination of these variables can result in differences as high as a factor of six between flow rates reported for a given boiler size class. Thus, it can be shown that reasonable design assumptions based on the literature lead to justification of the data as reported in the NEDS.

In order to prepare SO_2 control alternative cost calculations as in Parts II and III of this report, it was decided that flue gas flow rate data would be supplied by utilizing an alternative source of actual stack data. This was accomplished by a stack test data literature review for the specific sizes of boilers to be controlled. The size of boiler to be controlled was based on the SO_2 emissions calculations as described in the next section.

The flue gas flow rates were based on averages from Reference (5) and were as shown in Table 7.

TABLE 7. AVERAGE FLUE GAS FLOW RATE FOR VARIOUS BOILER SIZES AND FUELS

Fuel	Size (10^6 Btu/hr)	Flow Rate (scfm)
High Sulfur Oil	2	450
High Sulfur Oil	20	4430
High Sulfur Oil	250	63333
High Sulfur Pulverized Coal	20	5300
High Sulfur Pulverized Coal	250	62500
High Sulfur Stoker Coal	20	6660
High Sulfur Stoker Coal	250	68750

BOILER POPULATION CHARACTERIZATION

Table 8 presents data on the number, total installed capacity, and average hourly fuel use of boilers in each class defined by size, fuel type, and use category. Since the NEDS contains information on all sizes of boilers in the national inventory, the pertinent information for boilers other than non-utility combustion sources is included in the table. The information presented in the table is adjusted for all available data sources, including NEDS, ABMA sales data, and the boiler literature review (per Reference 6).

In dealing with SO_x control alternative analysis, the average hourly fuel use is the most important parameter shown. Average hourly fuel use is shown as the heat input equivalent of the amount of fuel used per hour by the total number of boilers in each class. It takes into account the annual load factor of each boiler class and is the parameter through which one may get an estimate of the actual amount of SO_x emissions by each boiler class in a year.

It is for the above reason that Figures 3 through 10 are shown. These figures are a graphical representation of average hourly fuel use by boiler size class and use category for each type of fuel considered in this study. In viewing these graphs, it is readily apparent that the boilers burning natural gas and by-product gas use more fuel per hour (heat equivalent) than other types of boilers. Oil fuel is second in quantity used, and coal is third.

By using the low/high sulfur content (Table 7) of each fuel and splitting the totals in Table 8 into their low/high sulfur components using a ratio from the NEDS data, it was possible to calculate SO_2 emissions from each class of boiler. These emissions were calculated using EPA emissions factors from Reference (8) and can be considered to be potential emissions. They are emissions which would result if each boiler in the country were fired at 100 percent of its design capacity for 24 hours a day for an entire year. In order to obtain an estimate of actual SO_2

TABLE 8. ESTIMATED NUMBER, TOTAL INSTALLED DESIGN CAPACITY (10^6 BTU/HR), AND AVERAGE HOURLY FUEL USE (10^6 BTU/HR) AS A FUNCTION OF DESIGN CAPACITY RANGE, FUEL TYPE, AND USE

	Class 1*: 0.5-1.0x10^6 Btu/hr			Class 1: 1^+ -2x10^6 Btu/hr			Class 2: 2^+ -5x10^6Btu/hr			Class 3: 5^+ -10x10^6Btu/hr			Class 4: 1^+ -2x10^7Btu/hr		
	Com.	Ind.	Ut.	Com.	Ind.	Ut.	Com.	Ind.	Ut.	Com.	Ind.	Ut.	Com.	Ind.	Ut.
Stoker coal	4,910 4,910 1,498	1,734 1,734 747	1 1 0	3,387 6,774 2,066	1,967 3,934 1,675	1 2 1	4,998 20,166 6,151	3,946 17,473 7,443	6 26 12	3,917 32,699 9,973	3,264 25,789 10,986	3 17 8	1,378 21,441 6,540	2,004 32,411 13,807	14 244 117
Pulverized coal	0 0 0	281 281 147	0 0 0	109 218 92	765 1,536 802	0 0 0	0 0 0	689 2,568 1,346	0 0 0	27 190 81	326 2,584 1,354	0 0 0	63 908 385	428 6,931 3,632	2 28 12
Residual oil	5,261 5,261 1,269	8,067 8,067 2,959	4 4 1	3,496 6,992 1,713	8,959 17,918 6,594	7 14 5	7,891 30,437 7,457	23,109 92,313 33,971	7 32 12	3,972 32,345 7,925	13,983 111,806 41,145	12 82 30	2,620 40,564 9,938	7,453 114,311 42,066	17 282 104
Distillate oil	19,992 19,992 4,118	14,169 14,169 4,676	2 2 0	13,547 27,094 5,581	15,733 31,466 10,384	4 8 1	11,461 43,401 8,941	16,789 65,320 21,556	14 56 8	3,047 22,824 4,702	4,758 37,514 12,380	5 36 5	856 11,879 2,447	1,764 26,879 8,870	5 90 12
By-Product Gas	0 0 0	421 421 265	0 0 0	0 0 0	874 1,748 1,101	0 0 0	0 0 0	877 3,064 1,933	0 0 0	0 0 0	245 2,176 1,371	0 0 0	0 0 0	84 1,315 828	0 0 0
Natural gas	11,644 11,644 3,703	15,853 15,853 8,212	1 1 0	7,866 15,732 5,003	17,590 35,180 18,223	2 4 2	6,137 24,174 7,687	27,118 106,591 55,214	9 35 17	2,992 24,157 7,682	12,976 102,013 52,843	11 78 37	981 15,355 4,883	6,493 100,021 51,811	33 489 232
Sum	41,807 41,807 10,608	40,545 40,545 17,016	8 8 1	28,405 56,810 14,455	45,888 91,782 38,780	14 28 9	30,487 118,178 30,236	72,528 287,333 121,463	36 151 49	13,955 112,215 30,363	35,582 281,882 120,079	31 213 80	5,898 90,147 24,193	18,226 281,868 121,014	71 1,133 477
Class sum		82,360 82,360 27,625			74,307 148,620 53,244			103,051 405,662 151,748			49,568 394,310 150,522			24,195 373,148 145,634	

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*Top number is number of boilers, middle number is total installed design capacity, and the bottom number is the product of annual load factor and total installed design capacity.

TABLE 8. ESTIMATED NUMBER, TOTAL INSTALLED DESIGN CAPACITY (10^6 BTU/HR), AND AVERAGE HOURLY FUEL USE (10^6 BTU/HR) AS A FUNCTION OF DESIGN CAPACITY RANGE, FUEL TYPE, AND USE (Continued)

	Class 5*: 2^+ $-5x10^7$ Btu/hr			Class 6: 5^+ $-10x10^7$ Btu/hr			Class 7: 1^+ $-2x10^8$ Btu/hr			Class 8: 2^+ $-5x10^8$ Btu/hr			Class 9: 5^+ $-10x10^8$ Btu/hr		
	Com.	Ind.	Ut.	Com.	Ind.	Ut.	Com.	Ind.	Ut.	Com.	Ind.	Ut.	Com.	Ind.	Ut.
Stoker coal	1,053	2,579	39	336	1,573	67	47	643	117	13	157	58	0	15	5
	35,447	91,957	1,482	24,609	115,979	5,546	6,088	92,010	17,701	3,831	45,085	16,503	0	11,071	3,005
	10,811	39,165	710	7,506	49,194	2,657	1,857	39,196	829	1,168	19,206	7,905	0	2,967	1,439
Pulverized coal	100	368	14	33	294	19	16	409	81	0	220	302	0	51	235
	3,289	14,100	592	2,518	24,679	1,549	2,229	60,155	12,517	0	66,569	103,283	0	33,022	168,744
	1,395	7,388	250	1,068	12,932	655	945	31,521	5,295	0	34,882	43,688	0	19,152	71,379
Residual oil	1,926	5,058	17	621	1,785	82	151	743	168	41	285	243	4	44	163
	65,042	169,137	282	44,009	131,806	5,497	20,914	106,429	25,930	13,009	86,684	74,605	4,209	30,273	111,307
	1,593	62,242	104	10,782	48,505	2,023	5,124	39,166	9,542	3,187	31,900	27,455	1,031	11,140	57,165
Distillate oil	442	858	5	121	203	14	41	114	43	17	29	67	3	3	33
	13,558	28,063	90	8,879	15,992	1,070	6,149	16,388	6,239	5,833	8,717	20,949	1,907	2,492	21,735
	2,793	9,261	12	1,829	5,264	139	1,267	5,408	811	1,202	2,877	2,723	393	822	2,826
By-Product Gas	0	174	0	0	170	0	0	137	0	0	60	4	0	26	0
	0	7,038	0	0	12,124	0	0	18,108	0	0	19,181	1,496	0	16,554	0
	0	4,447	0	0	7,638	0	0	11,408	0	0	12,069	942	0	10,429	0
Natural gas	684	5,000	56	224	2,158	43	100	918	181	36	324	33	5	75	241
	21,721	276,842	1,950	17,691	156,194	6,161	13,982	136,580	27,733	12,099	98,373	108,281	2,794	48,744	169,221
	6,967	149,404	924	5,626	80,908	2,920	4,319	70,798	13,145	3,835	50,957	51,325	885	25,249	8,021
Sum	4,205	14,037	131	1,335	6,183	265	355	2,964	590	107	1,075	707	12	214	677
	139,057	587,137	4,396	98,706	456,774	19,823	49,362	429,670	90,118	34,772	324,609	325,118	8,910	142,161	474,012
	23,559	271,907	2,000	26,811	204,441	8,394	13,512	197,497	29,622	9,392	151,911	134,038	2,312	69,759	140,830
Class sum		18,373			7,783			3,909			1,889			903	
		730,590			575,303			569,150			684,499			625,083	
		297,466			239,646			240,631			295,341			212,901	

* Top number is number of boilers, middle number is total installed design capacity, and the bottom number is the product of annual load factor and total installed design capacity.

TABLE 8. ESTIMATED NUMBER, TOTAL INSTALLED DESIGN CAPACITY (10^6 BTU/HR), AND AVERAGE HOURLY FUEL USE (10^6 BTU/HR) AS A FUNCTION OF DESIGN CAPACITY RANGE, FUEL TYPE, AND USE (Continued)

	Class 10*: $1^+-2\times10^9$ Btu/hr			Class 11: $2^+-5\times10^9$ Btu/hr			Class 12: $5^+-10\times10^9$ Btu/hr			Class 13: $1^+-2\times10^{10}$ Btu/hr			Class 14: > 2×10^{10} Btu/hr		
	Com.	Ind.	Ut.	Com.	Ind.	Ut.	Com.	Ind.	Ut.	Com.	Ind.	Ut.	Com.	Ind.	Ut.
Stoker coal	0 0 0	5 6,851 1,836	2 3,123 1,496	0 0 0	2 7,600 2,037	2 6,132 2,937	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0
Pulverized coal	0 0 0	10 16,019 9,291	256 370,857 159,410	0 0 0	5 11,995 6,928	125 378,536 160,121	0 0 0	1 5,068 2,939	53 327,022 138,330	0 0 0	0 0 0	2 22,466 9,503	0 0 0	0 0 0	4 - -
Residual oil	0 0 0	12 17,450 6,422	83 114,113 57,910	0 0 0	1 3,020 1,111	43 128,886 70,068	0 0 0	0 0 0	5 26,906 17,408	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0
Distillate oil	0 0 0	0 0 0	3 4,129 537	0 0 0	1 2,274 750	1 2,854 371	0 0 0	0 0 0	1 5,184 674	0 0 0	0 0 0	1 13,070 1,699	0 0 0	0 0 0	0 0 0
By-Product Gas	0 0 0	4 5,604 3,531	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0	0 0 0
Natural gas	4 6,301 2,004	32 46,063 23,861	174 242,621 115,010	1 2,626 835	11 28,908 14,474	80 247,992 117,548	0 0 0	5 38,964 20,183	17 101,915 48,308	1 10,350 3,291	4 61,550 31,883	1 10,931 5,191	1 - -	2 - -	1 - -
Sum	4 6,301 2,004	63 91,987 44,941	528 734,837 354,404	1 2,626 835	20 53,797 25,300	251 764,400 351,045	0 0 0	6 44,032 23,122	76 461,027 204,720	1 10,350 3,291	4 61,550 31,883	4 46,517 16,393	1 - -	2 - -	5 - -
Class sum		595 833,125 401,349			272 820,823 377,180			82 505,059 227,842			9 118,417 51,567			8 - -	

* Top number is number of boilers, middle number is total installed design capacity, and the bottom number is the product of annual load factor and total installed design capacity.

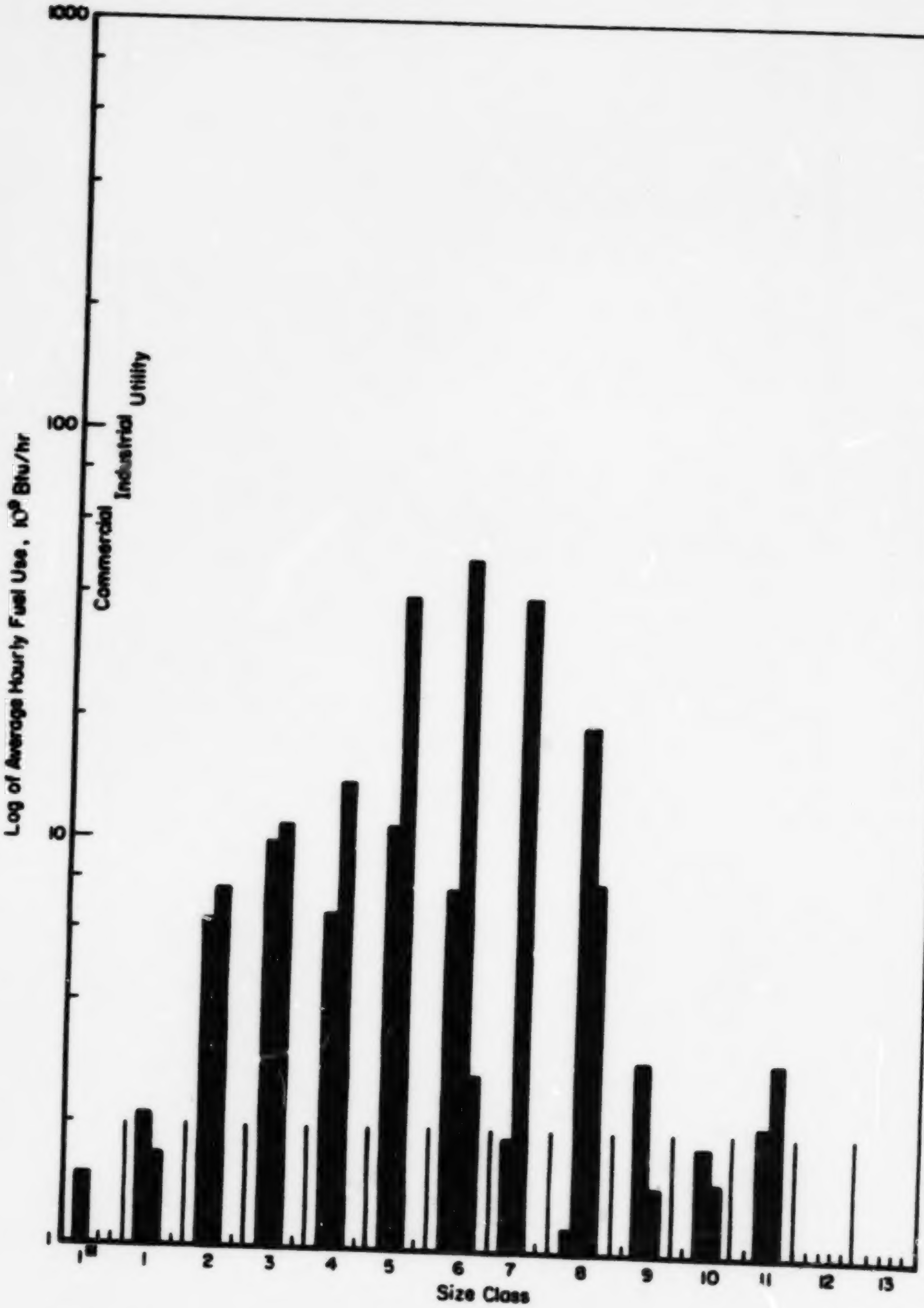


FIGURE 3. ANNUAL HOURLY FUEL USE BY SIZE CLASS AND USE CATEGORY FOR STOKER COAL FUEL

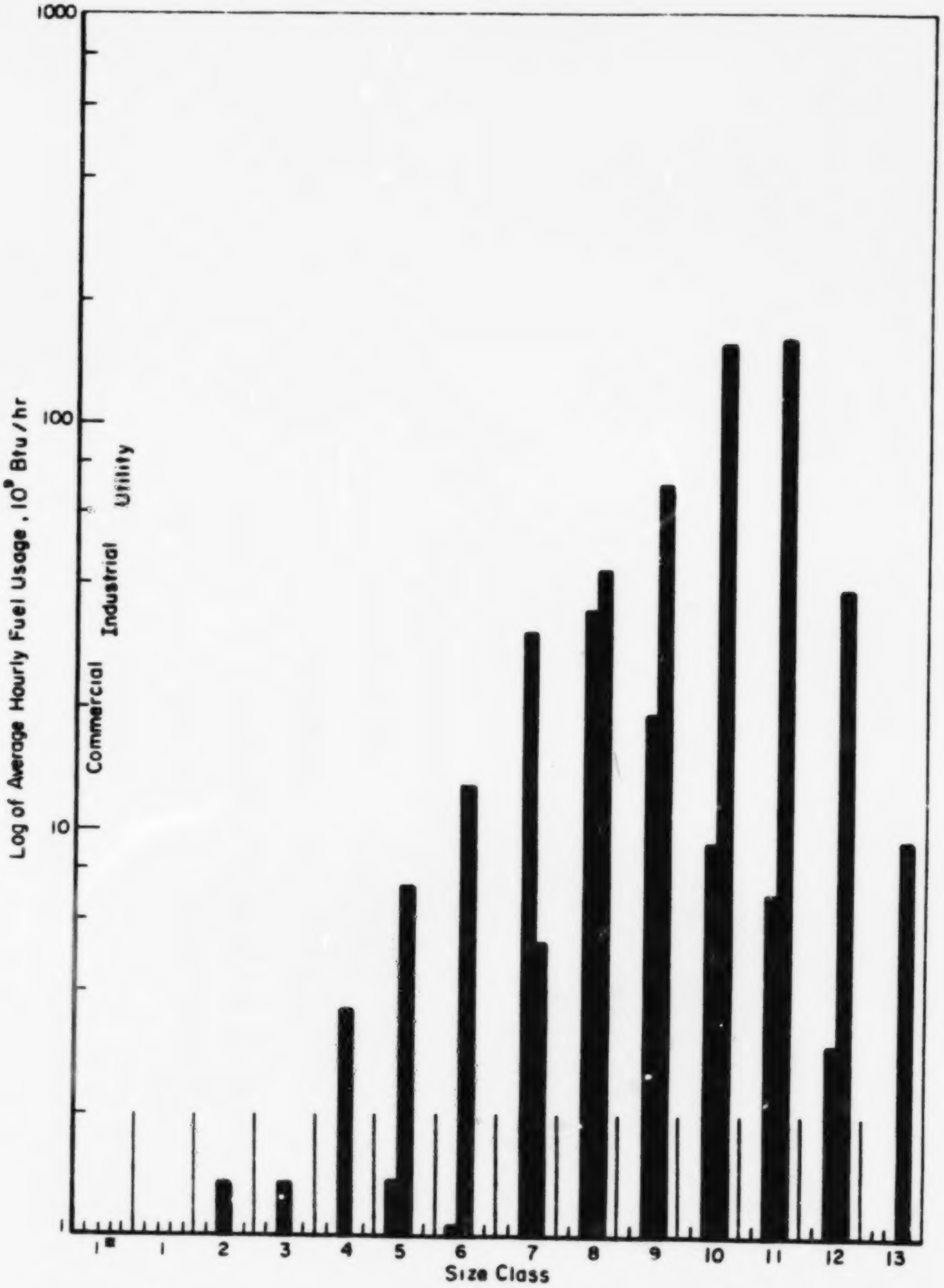


FIGURE 4. ANNUAL HOURLY FUEL USAGE (10^9 Btu/hr) BY SIZE CLASS AND USE CATEGORY FOR PULVERIZED COAL FUEL

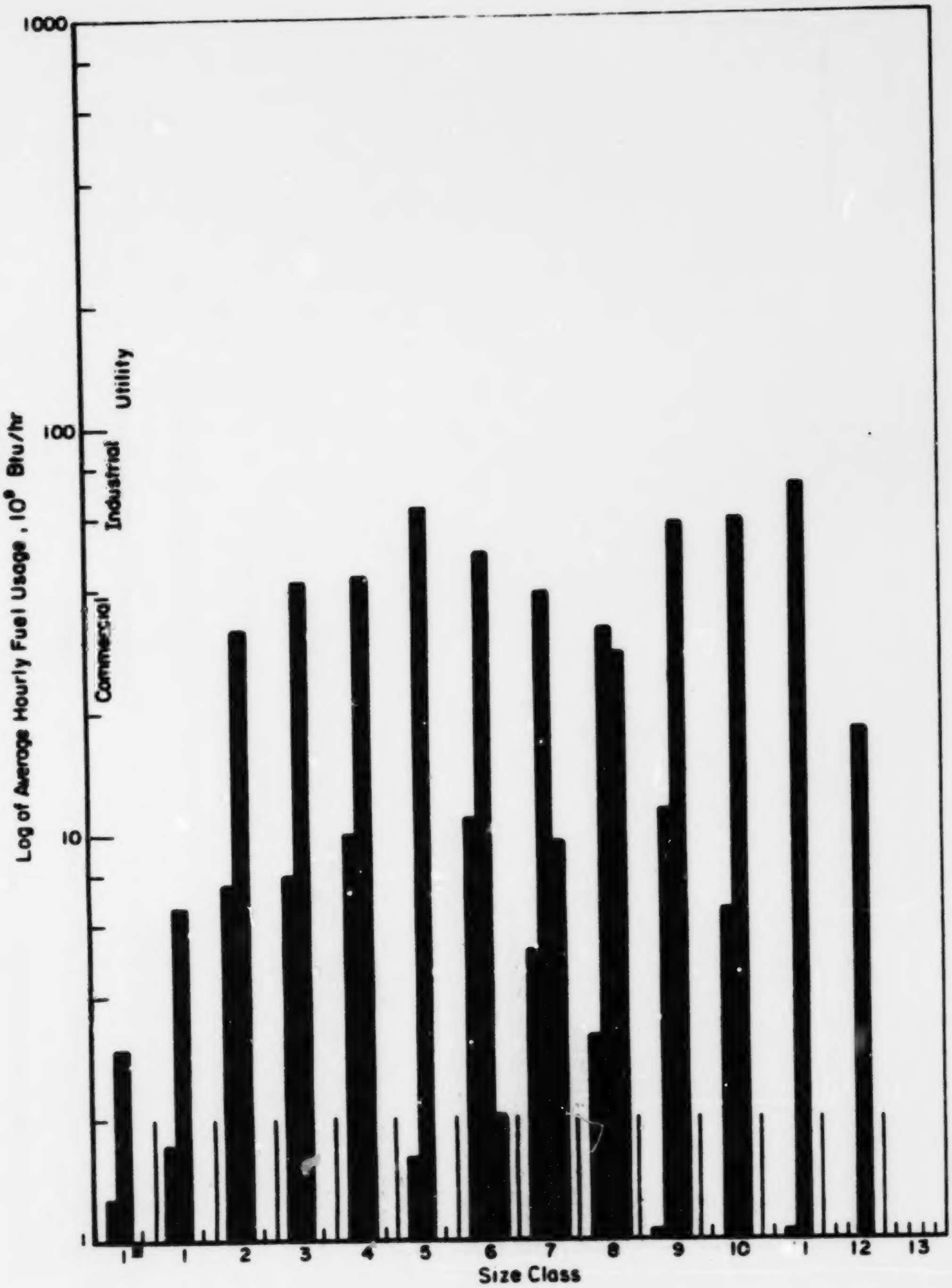


FIGURE 5. ANNUAL HOURLY FUEL USE BY SIZE CLASS AND USE CATEGORY FOR RESIDUAL OIL FUEL

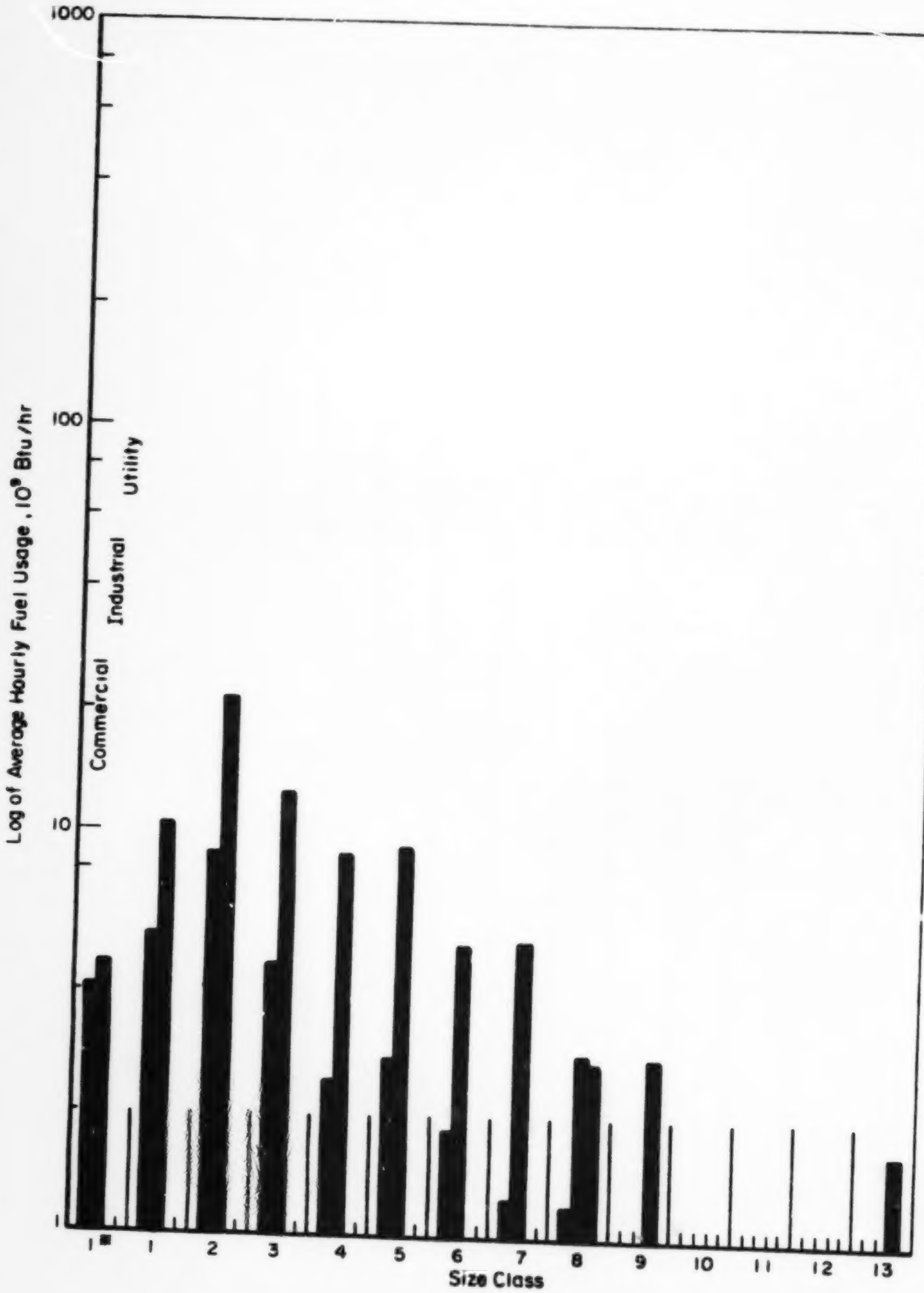


FIGURE 6. ANNUAL HOURLY FUEL USE BY SIZE CLASS AND USE CATEGORY FOR DISTILLATE OIL FUEL

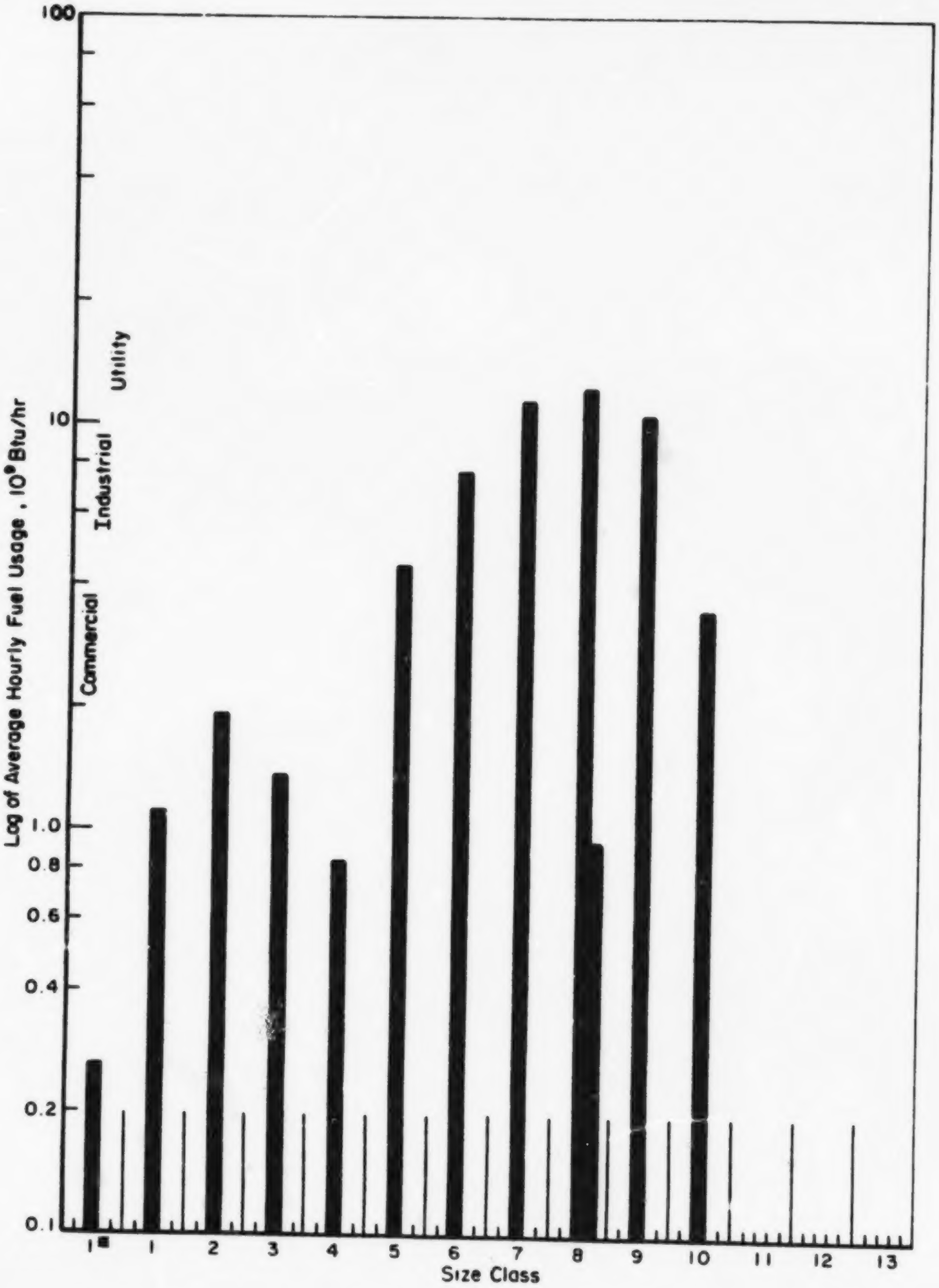


FIGURE 7. ANNUAL HOURLY FUEL USE BY SIZE CLASS AND USE CATEGORY FOR BY-PRODUCT GAS FUEL

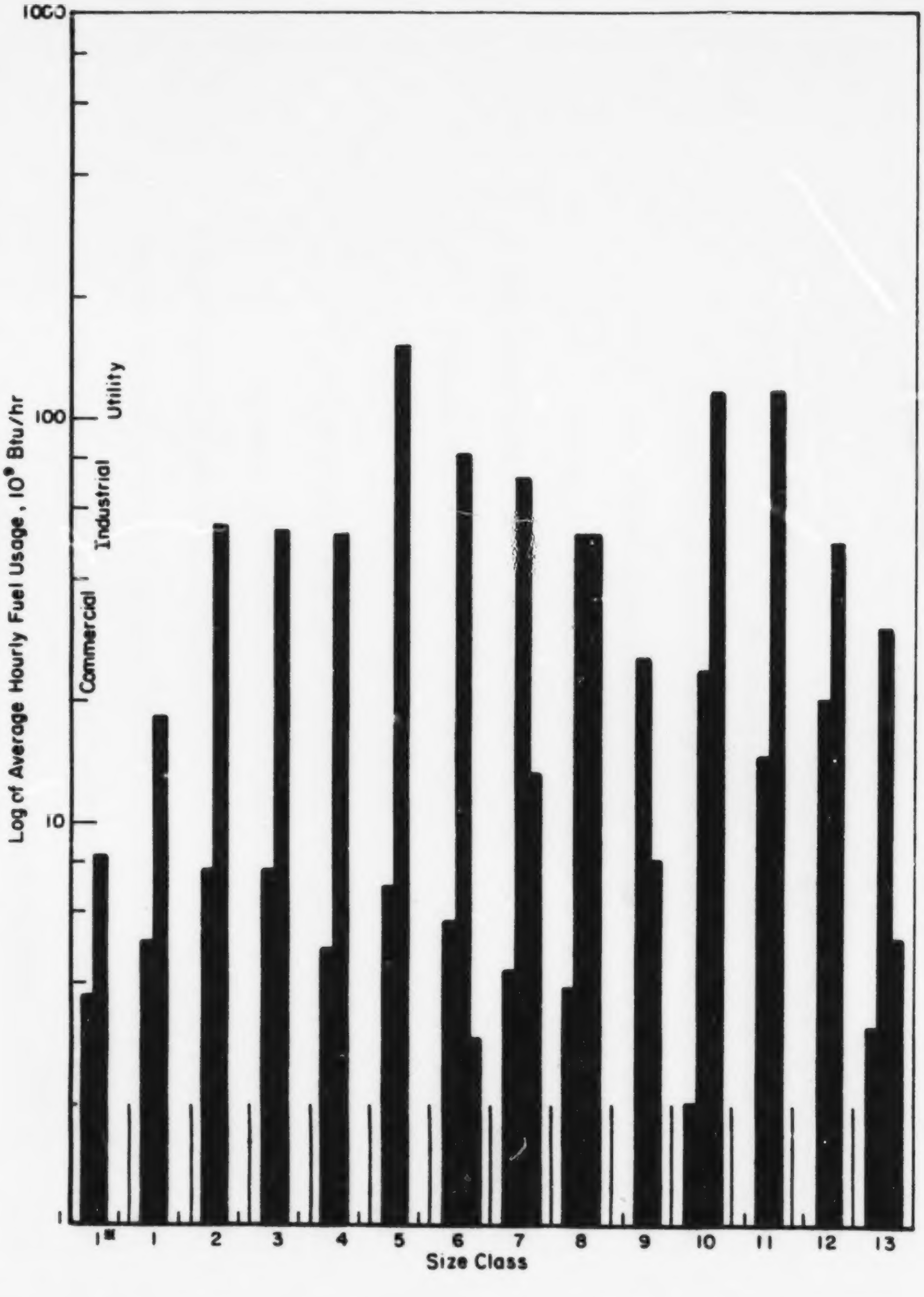


FIGURE 8. ANNUAL HOURLY FUEL USE BY SIZE CLASS AND USE CATEGORY FOR NATURAL GAS FUEL

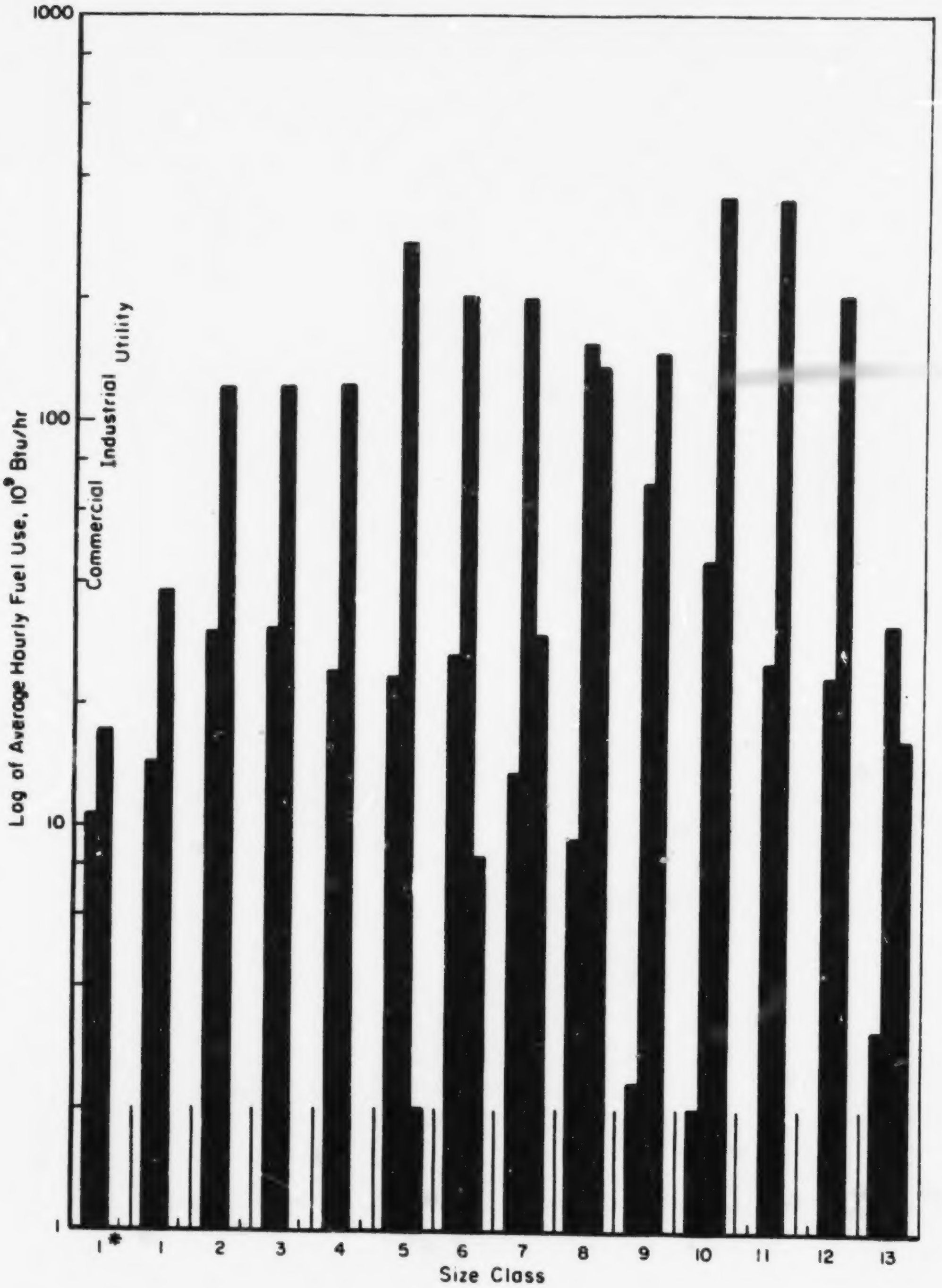


FIGURE 9. ANNUAL HOURLY FUEL USE BY SIZE CLASS AND USE CATEGORY FOR THE TOTAL OF ALL FUELS PER SIZE CLASS

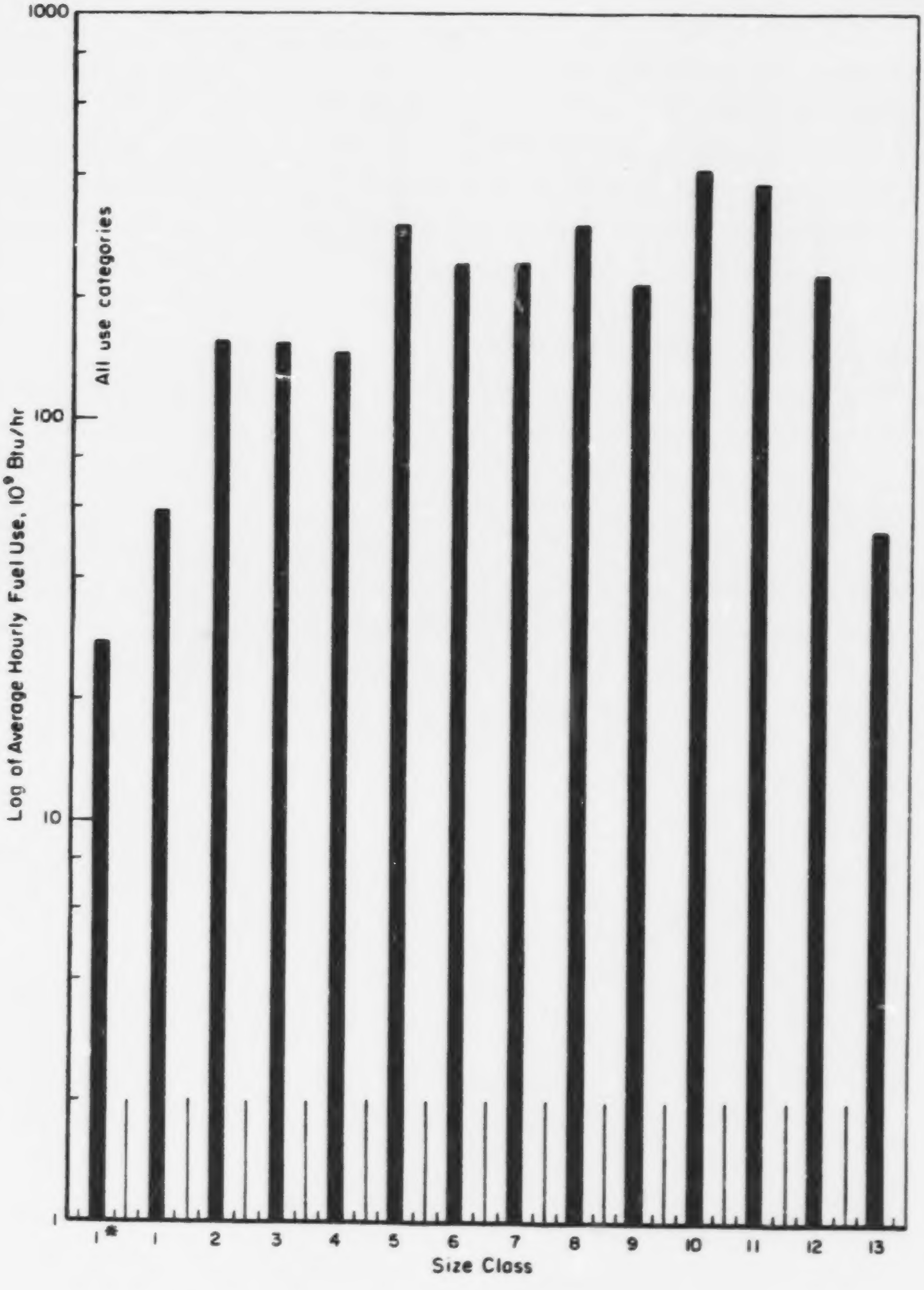


FIGURE 10. ANNUAL HOURLY FUEL USE BY SIZE CLASS FOR ALL USE CATEGORIES AND ALL FUELS IN EACH CLASS

emissions, one can adjust the boiler design capacity in each class by its appropriate annual load factor. The figures in Table 8 were adjusted by load factor in calculating the average hourly fuel use. Applying EPA emission factors to the average hourly fuel use (from Table 8) values will thus yield an estimate of actual SO_2 emission for each class. Table 9 (from Reference 6) shows the potential and actual SO_2 emissions for each class of boiler as calculated in the aforementioned manner. Once again, in addressing SO_2 control alternative analysis, the estimate of actual SO_2 emissions is the important parameter. Figures 11 through 20 are graphical illustrations of estimated actual SO_2 emissions for each size class, use category, and fuel. By-product gas and high sulfur natural gas are not shown since their emissions were negligible.

The graphs readily show that preponderance of non-utility SO_2 emissions appears to come from high-sulfur coal-fired units in classes 5 through 8 (10×10^6 Btu/hr to 500×10^6 Btu/hr) and from high-sulfur oil-fired units in classes 1 through 8 (1×10^6 Btu/hr through 500×10^6 Btu/hr).

TABLE 9. SULFUR DIOXIDE EMISSIONS (TONS/YR) BY BOILER SIZE CLASS$^{(1)}$

Fuel	Class 1*: 5^+-10 x 10^5 Btu/hr			Class 1: 1^+-2 x 10^6 Btu/hr			Class 2: 2^+- 5 x 10^6 Btu/hr			Class 3: 5^+-10 x 10^6 Btu/hr		
	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility
Stoker Coal, Low Sulfur	13,993.0	5,027.5	0	19,305.3	8,143.8	0	55,398.3	57,938.	0	186,993.2	74,815.7	0
	4,269.3	2,509.3	0	5,887.9	4,065.8	0	16,897.5	24,681.8	0	57,032.1	31,871.7	0
Stoker Coal, High Sulfur	40,155.0	3,497.2	15.6	55,398.0	12,746.1	31.3	169,298.4	501,124.2	1,941.9	154,839.3	313,632.4	0
	12,250.0	1,489.1	0	16,896.1	5,430.2	15.6	31,638.9	213,466.7	896.2	47,224.6	133,605.8	0
Pulverized Coal, Low Fulfur	0	944.8	0	821.2	5,164.4	0	0	8,378.6	0	735.0	11,718.2	0
	0	494.3	0	346.5	2,696.5	0	0	4,496.2	0	313.4	6,140.3	0
Pulverized Coal, High Sulfur	0	3,118.0	0	0	17,063.5	0	0	27,816.6	0	0	0	0
	0	1,087.4	0	0	8,899.0	0	0	14,580.8	0	0	0	0
Residual Oil, Low Sulfur	12,299.0	27,378.0	0	16,345.4	60,836.2	40.9	92,259.0	354,569.5	211.4	63,543.9	302,675.7	344.5
	2,394.8	10,080.0	0	3,182.6	22,388.1	14.7	22,603.5	130,473.5	79.2	15,569.0	11,135.7	199.3
Residual Oil, High Sulfur	39,770.0	78,760.0	39.2	52,855.4	174,938.2	19.6	85,125.3	1,050,932.6	52.4	1,074,148.4	1,245,368.0	165.5
	9,744.3	28,987.0	9.9	12,949.5	64,379.7	6.9	2,055.4	386,741.1	19.8	263,184.3	458,300.0	60.4
Distillate Oil, Low Sulfur	39,001.0	17,713.0	0.9	35,237.4	39,336.0	3.4	49,072.0	60,079.7	34.0	33,491.2	92,901.1	15.7
	5,355.6	5,865.5	0	7,258.4	12,981.2	.4	9,834.9	12,377.0	4.9	6,899.6	30,658.3	2.2
Distillate Oil, High Sulfur	12,555.3	4,301.3	0	17,015.5	9,552.4	0	26,593.3	24,484.6	0	12,684.5	31,636.6	0
	2,586.3	1,419.6	0	3,504.9	3,152.6	0	5,478.3	5,044.3	0	2,613.4	10,447.0	0
By-Product Gas Low Sulfur	0	9.8	0	0	40.5	0	0	71.0	0	0	41.8	0
	0	6.1	0	0	25.5	0	0	44.8	0	0	26.4	0
By-Product Gas High Sulfur	0	0	0	0	0	0	0	0	0	0	714.3	0
	0	0	0	0	0	0	0	0	0	0	135.0	0
Natural Gas, Low Sulfur	30.6	40.9	0	41.4	92.5	0	63.6	280.3	0	63.5	768.3	0
	9.7	21.6	0	13.2	47.9	0	20.2	145.2	0	20.2	139.0	0
Natural Gas, High Sulfur	0	0	0	0	0	0	0	0	0	0	0	0
	0	0	0	0	0	0	0	0	0	0	0	0

(1) 1* indicates either model fuel sulfur content of zero or no installed capacity in the category.

TABLE 9. ESTIMATED SULFUR DIOXIDE EMISSIONS (TONS/YR) BY BOILER SIZE CLASS (Continued)

Fuel	Class 4: 1^+-2×10^7 Btu/hr			Class 5: 2^+-5×10^7 Btu/hr			Class 6: 5^+-10×10^7 Btu/hr			Class 7: 1^+-2×10^8 Btu/hr		
	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility
Stoker Coal, Low Sulfur	49,113.2 14,980.6	101,075.7 43,058.3	8,488.7 4,074.9	87,819.6 26,784.0	187,626.8 79,928.7	3,601.2 759.1	66,312.8 19,920.6	249,368.1 105,772.9	25,293.8 12,118.1	6,334.6 1,932.1	254,713.7 108,507.3	60,636.2 2,839.5
Stoker Coal, High Sulfur	206,802.6 64,869.7	561,292.9 238,886.4	1,980.1 948.8	330,131.8 100,687.8	1,429,050.3 608,772.8	24,912.2 11,934.8	257,673.4 78,593.9	1,502,051.4 637,114.0	52,515.3 25,158.5	94,402.6 28,795.5	1,092,888.4 465,567.5	170,690.9 7,994.5
Pulverized Coal, Low Sulfur	0 0	36,825.1 19,297.3	9.3 40.0	14,272.5 6,053.5	27,462.6 14,389.6	1,459.4 616.0	3,816.2 1,618.5	37,197.8 19,869.5	1,724.4 729.5	7,356.2 3,118.9	130,612.4 68,440.7	11,133.0 4,709.9
Pulverized Coal, High Sulfur	5,727. 2,428.3	32,045.2 16,791.9	222.3 95.3	22,297.2 9,457.4	179,053.5 93,818.9	1,993.6 841.8	11,362.4 4,823.3	314,961.7 165,041.7	42,726.6 18,068.6	10,351.7 4,472.9	699,090.0 366,320.	264,865.6 117,045.8
Residual Oil, Low Sulfur	79,171.9 19,396.6	295,755.5 108,100.6	455.7 168.9	107,708.6 2,637.9	333,207.6 122,619.7	290.9 107.3	78,357.5 19,197.2	321,849.8 11,844.2	7,562.3 2,783.1	44,111.5 10,880.9	244,067.9 89,897.2	64,851.2 23,864.6
Residual Oil, High Sulfur	316,506.0 77,542.9	839,968.0 309,105.	3,320.4 1,225.1	414,158.7 10,143.8	1,300,827. 478,701.8	1,949.6 719.0	288,704.7 70,731.1	1,099,533. 404,632.	39,487.3 14,532.4	116,918.4 28,646.5	739,378.5 272,092.4	130,721.8 48,104.4
Distillate Oil, Low Sulfur	9,435.5 1,943.6	38,537.8 12,717.4	83.2 11.1	11,580.6 2,385.7	26,111.3 8,616.9	0 0	9,110.2 1,876.7	33,281.4 10,955.	1,438.7 186.9	10,141.2 2,089.6	75,331.1 24,859.1	7,774.1 1,010.5
Distillate Oil, High Sulfur	15,086.6 3,107.8	8,811.6 2,907.6	0 0	11,001.8 2,266.2	48,802.5 16,105.4	0 0	3,499.1 720.6	32,331.1 10,642.3	4,304.5 558.2	2,017.2 415.6	6,358.5 2,098.3	0 0
By-Product Gas, Low Sulfur	0 0	30.5 19.2	0 0	0 0	106.3 67.0	0 0	0 0	212.8 134.1	0 0	0 0	314.6 198.2	0 0
By-Product Gas, High Sulfur	0 0	0 0	0 0	0 0	572.3 360.6	0 0	0 0	681.0 429.0	0 0	0 0	1,048.0 660.7	0 0
Natural Gas, Low Sulfur	40.4 12.8	263. 1,226.3	1.3 6.1	57.1 18.3	728. 392.9	5.2 2.4	46.5 14.8	410.8 212.8	16.2 7.7	36.8 11.4	359. 186.	72.9 34.6
Natural Gas, High Sulfur	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0

(1) 1* indicates either model fuel sulfur content of zero or no installed capacity in the category.

TABLE 9. ESTIMATED SULFUR DIOXIDE EMISSIONS (TONS/YR) BY BOILER SIZE CLASS (Continued)

Fuel	Class 8: 2^{+}-5 x 10^{8} Btu/hr			Class 9: 5^{+}-10 x 10^{8} Btu/hr			Class 10: 1^{+}-2 x 10^{9} Btu/hr		
	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility
Stoker Coal, Low Sulfur	6,927.6 2,111.9	86,723.6 36,943.9	3,199.8 1,532.7	0 0	6,169.6 1,653.5	1,892.3 5,694.9	0 0	27,157.6 7,277.9	0 0
Stoker Coal, High Sulfur	59,889.0 18,260.1	227,296.1 96,827.0	268,163.2 128,451.0	0 0	471,614.2 176,393.5	41,765.2 20,000.1	0 0	46,209.7 12,383.6	158,391.8 75,873.9
Pulverized Coal, Low Sulfur	0 0	99,358.1 52,063.5	99,744.3 42,191.4	0 0	41,274.4 73,938.1	233,517.5 98,778.0	0 0	27,751.7 16,096.0	493,300.2 212,041.2
Pulverized Coal, High Sulfur	0 0	716,952.4 373,681.0	2,896,232.3 1,225,072.4	0 0	1,568,542.4 909,683.0	4,439,993.3 1,878,124.8	0 0	253,371.6 146,954.9	9,026,003.0 3,879,758.0
Residual Oil, Low Sulfur	23,647.4 5,793.0	71,458.5 26,797.0	194,919.3 71,731.2	25,157.2 6,162.3	57,840.4 21,704.4	345,274.4 177,839.5	0 0	15,705.8 5,780.0	205,602.5 104,339.0
Residual Oil, High Sulfur	36,086.9 8,841.0	756,032.4 278,222.2	494,003.2 181,797.6	0 0	203,247.9 74,792.0	471,242.3 242,020.2	0 0	90,857.1 33,438.0	234,660.3 119,085.3
Distillate Oil, Low Sulfur	5,681.0 1,170.7	5,621.2 1,855.3	22,203.6 2,886.1	1,689.8 348.2	1,709.2 563.8	21,967.0 2,856.1	0 0	0 0	17,016.4 2,213.1
Distillate Oil, High Sulfur	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0
Non-Natural Gas, Low Sulfur	0 0	412.6 260.1	0 0	0 0	0 0	0 0	0 00	0 0	0 0
By-Product Gas , High Sulfur	0 0	1,110.9 700.2	693.2 436.5	0 0	3,835.1 2,416.1	0 0	0 0	1,298.3 818.0	0 0
Natural Gas, Low Sulfur	71.8 10.0	258.7 134.0	284.8 135.0	7.3 2.3	128.0 66.4	445.0 21.1	16.6 5.3	121.1 62.7	638.0 302.5
Natural Gas, High Sulfur	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0

(1) 1^{*} indicates either model fuel sulfur content or zero or no installed capacity in the category.

TABLE 9. ESTIMATED SULFUR DIOXIDE EMISSIONS (TONS/YR) BY BOILER SIZE CLASS (Continued)

Fuel	Class 11: 2^{+}-5×10^{9} Btu/hr			Class 12: 5^{+}-10×10^{9} Btu/hr			Class 13: 1^{+}-2×10^{10} Btu/hr		
	Com.	Ind.	Utility	Com.	Ind.	Utility	Com.	Ind.	Utility
Stoker Coal,	0	0	0	0	0	0	0	0	0
Low Sulfur	0	0	0	0	0	0	0	0	0
Stoker Coal,	0	177,243.8	157,332.4	0	0	0	0	0	0
High Sulfur	0	47,506.0	75,356.4	0	0	0	0	0	0
Pulverized Coal,	0	0	1,019,488.0	0	0	593,083.0	0	0	0
Low Sulfur	0	0	43,126.5	0	0	750,874.0	0	0	0
Pulverized Coal,	0	84,826.2	9,839,853.0	0	232,136.0	6,370.437.0	0	0	765,264.0
High Sulfur	0	48,993.4	4,162,264.0	0	134,613.0	2,694,689.0	0	0	323,703.0
Residual Oil,	0	0	407,225.9	0	0	0	0	0	0
Low Sulfur	0	0	221,385.3	0	0	0	0	0	0
Residual Oil,	0	0	716,123.3	0	0	306,257.5	0	0	0
High Sulfur	0	0	389,316.1	0	0	198,146.6	0	0	0
Distillate Oil,	0	0	4,539.0	0	0	16,124.9	0	0	22,804.9
Low Sulfur	0	0	590.0	0	0	2,096.5	0	0	2,964.5
Distillate Oil,	0	0	0	0	0	0	0	0	0
High Sulfur	0	0	0	0	0	0	0	0	0
By-Product Gas,	0	0	0	0	0	0	0	0	0
Low Sulfur	0	0	0	0	0	0	0	0	0
By-Product Gas,	0	0	0	0	0	0	0	0	0
High Sulfur	0	0	0	0	0	0	0	0	0
Natural Gas,	6.9	76.0	652.2	0	102.5	268.0	27.2	161.9	28.9
Low Sulfur	2.2	38.0	309.1	0	53.1	127.0	8.7	83.8	13.7
Natural Gas,	0	0	0	0	0	0	0	0	0
High Sulfur	0	0	0	0	0	0	0	0	0

(1) 1^{*} indicates either model fuel sulfur content of zero or no installed capacity in the category.

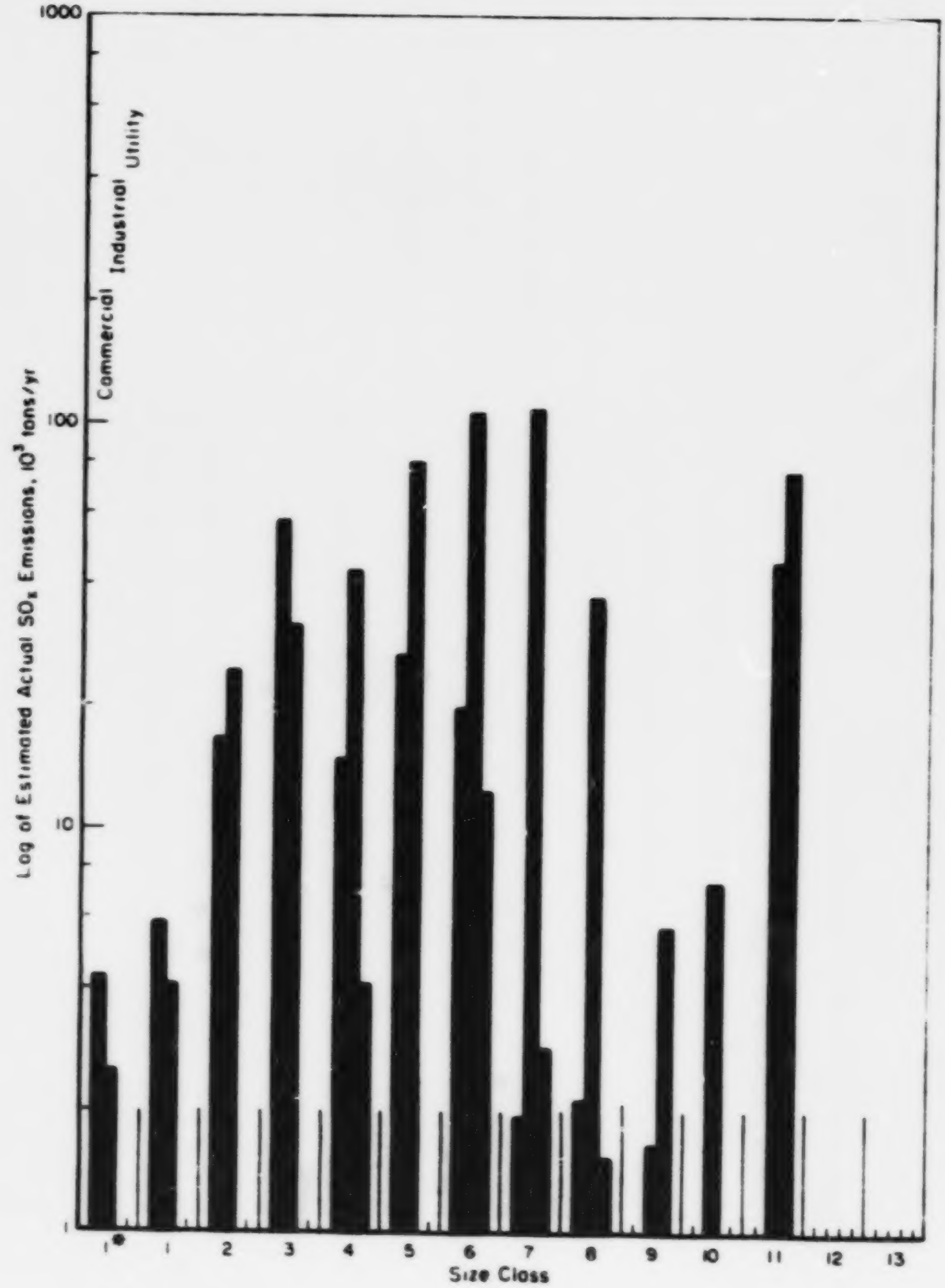


FIGURE 11. ESTIMATED ACTUAL SO$_x$ EMISSIONS BY SIZE CLASS AND USE CATEGORY FOR LOW SULFUR STOKER COAL FUEL

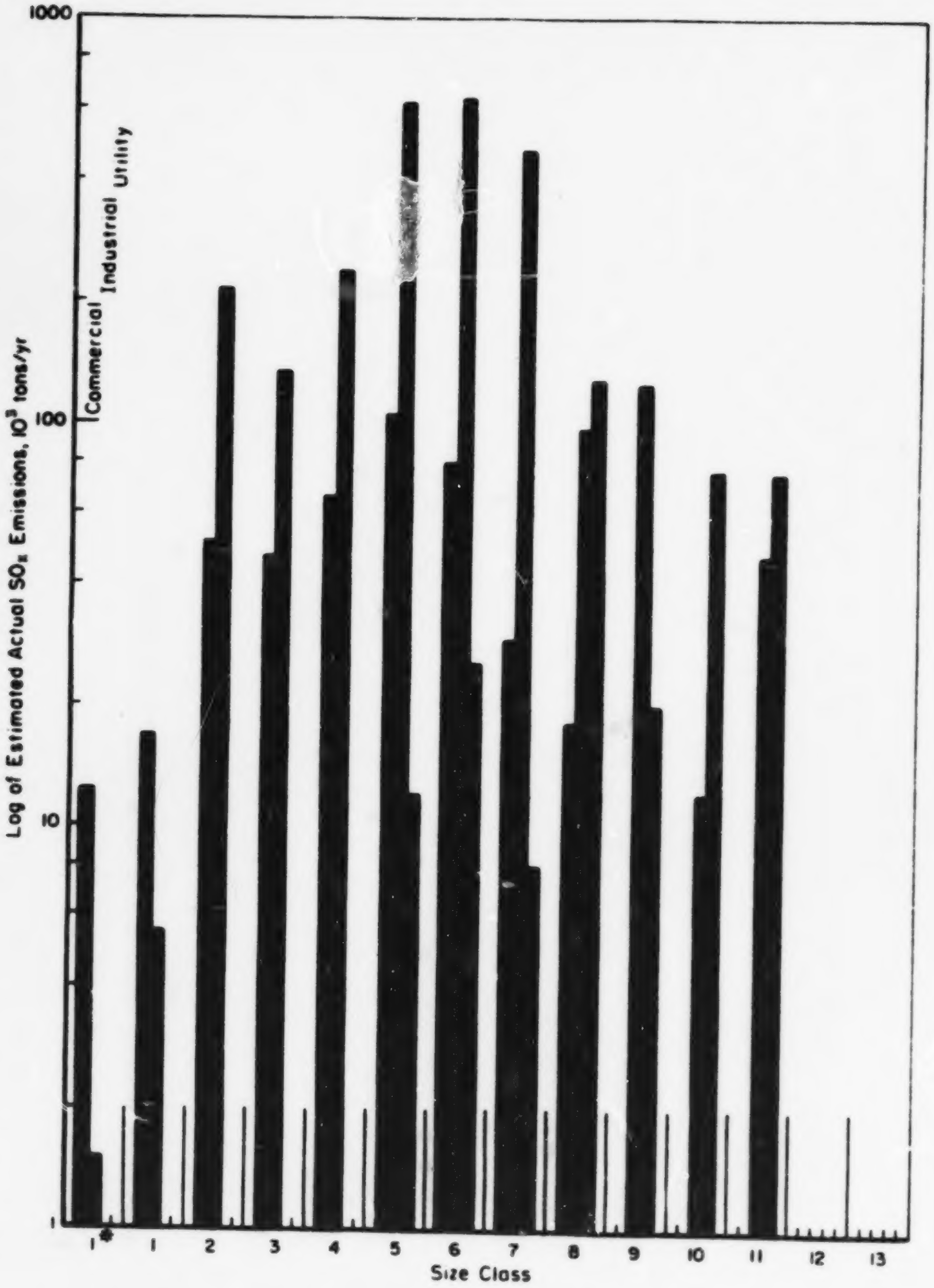


FIGURE 12. ANNUAL HOURLY FUEL USE (10^9 Btu/hr) BY SIZE AND USE CATEGORY FOR HIGH SULFUR STOKER COAL

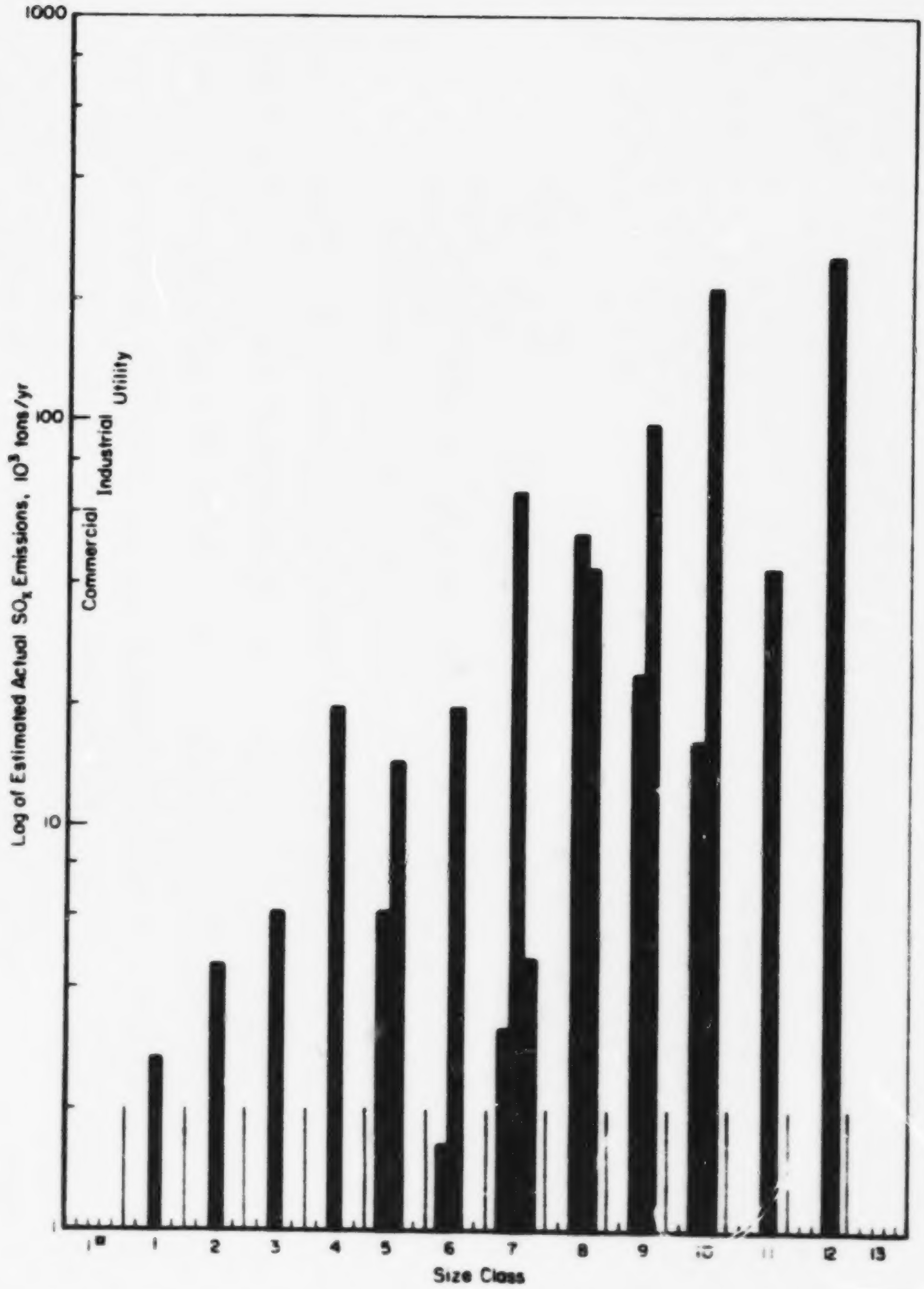


FIGURE 13. ESTIMATED ACTUAL SO_x EMISSIONS BY SIZE CLASS AND USE CATEGORY FOR LOW SULFUR PULVERIZED COAL FUEL

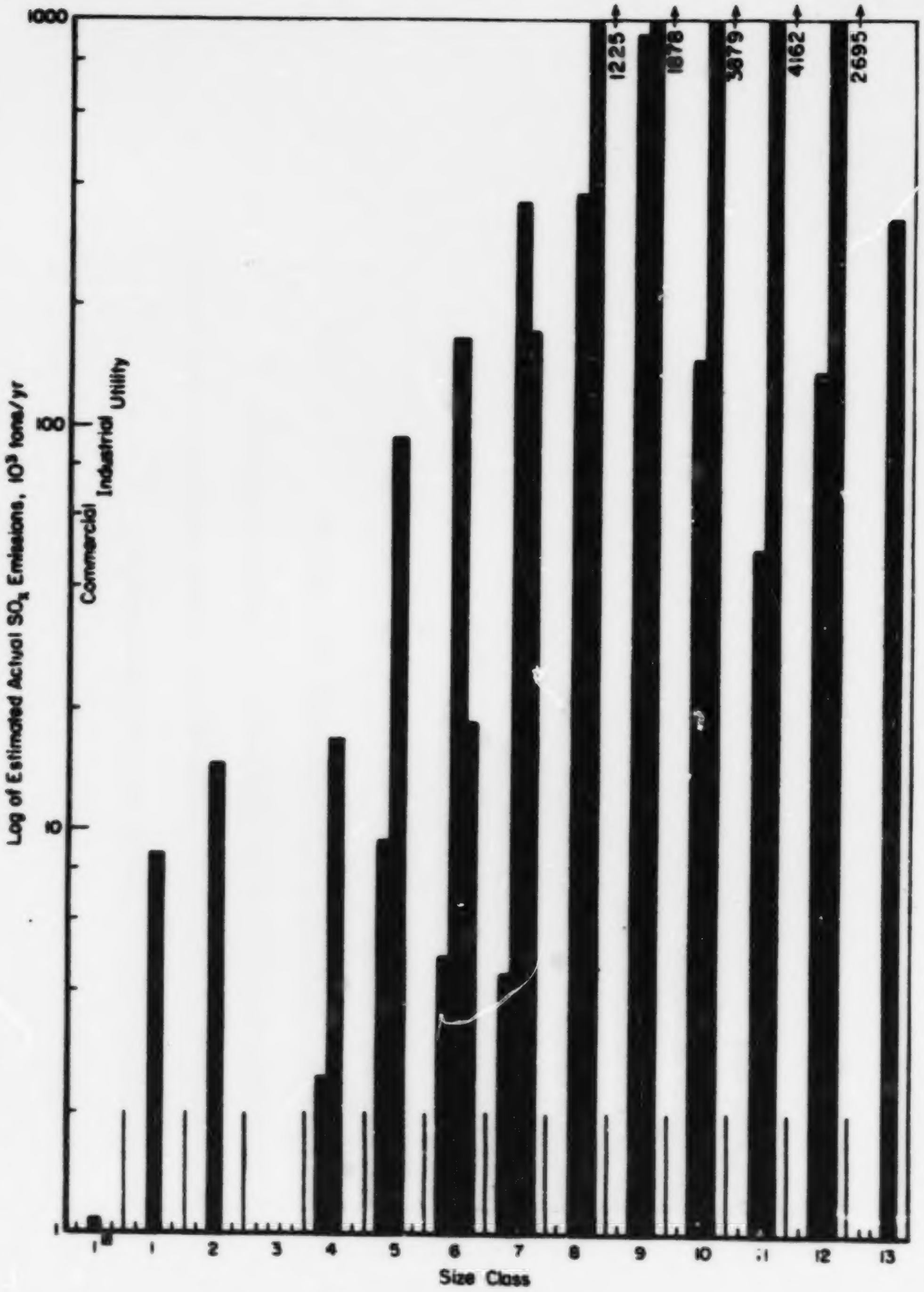


FIGURE 14. ANNUAL HOURLY FUEL USE (10^9 Btu/hr) BY SIZE AND USE CATEGORY FOR HIGH SULFUR PULVERIZED COAL FUEL

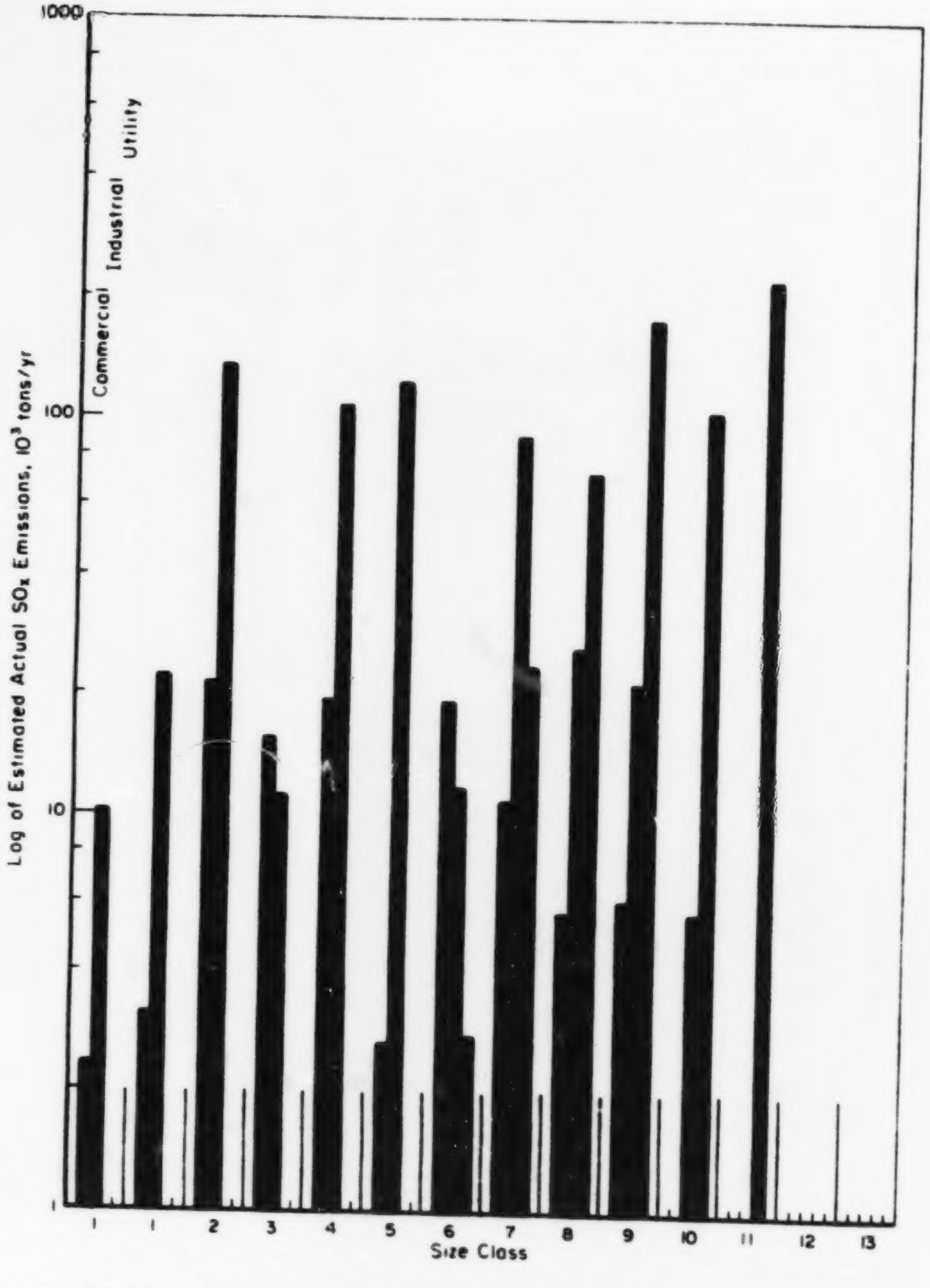


FIGURE 15. ESTIMATED ACTUAL SO_x EMISSIONS BY SIZE CLASS AND USE CATEGORY FOR LOW SULFUR RESIDUAL OIL FUEL

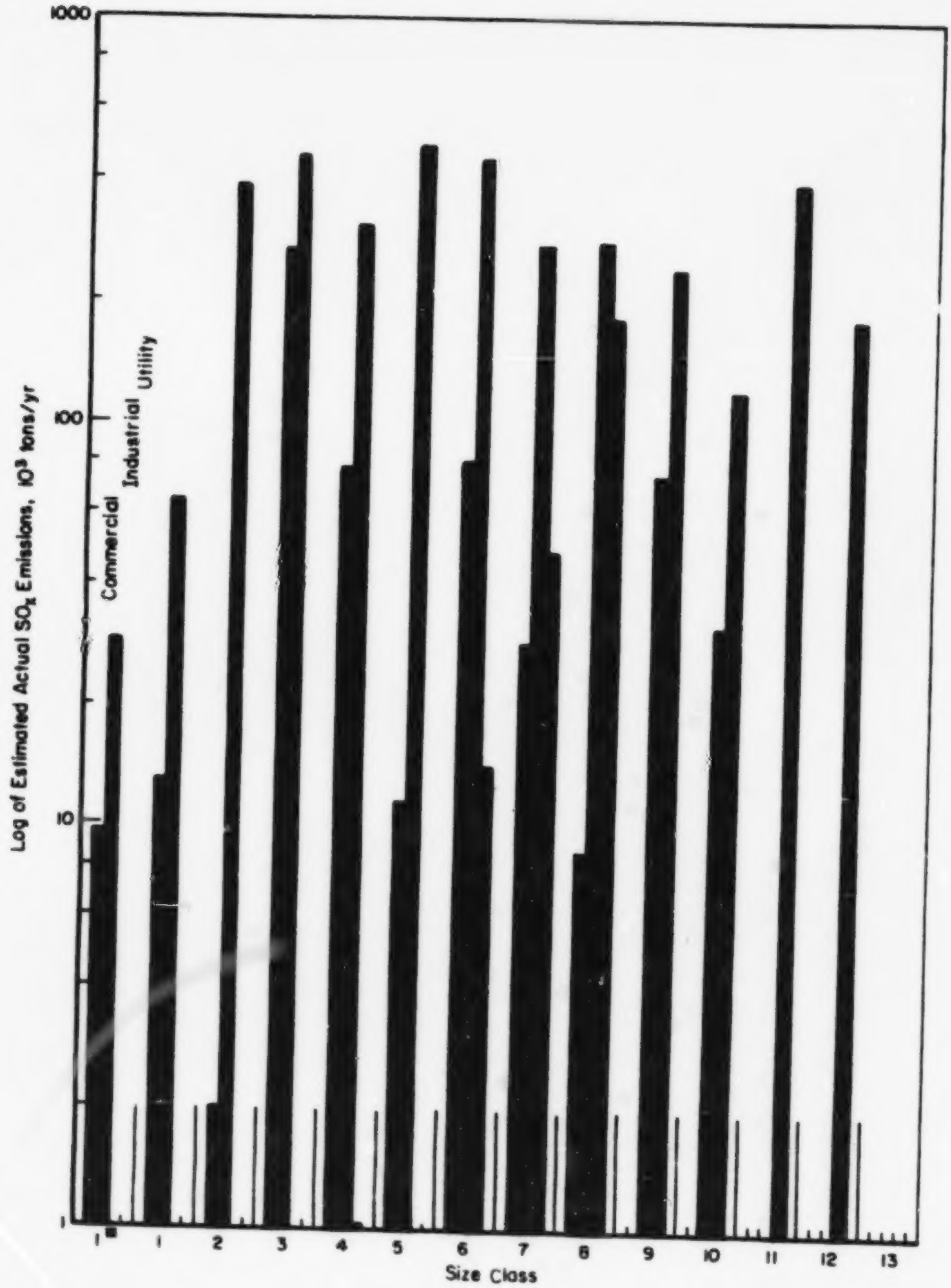


FIGURE 16. ANNUAL HOURLY FUEL USE (10^9 Btu/hr) BY SIZE AND USE CATEGORY FOR HIGH SULFUR RESIDUAL OIL FUEL

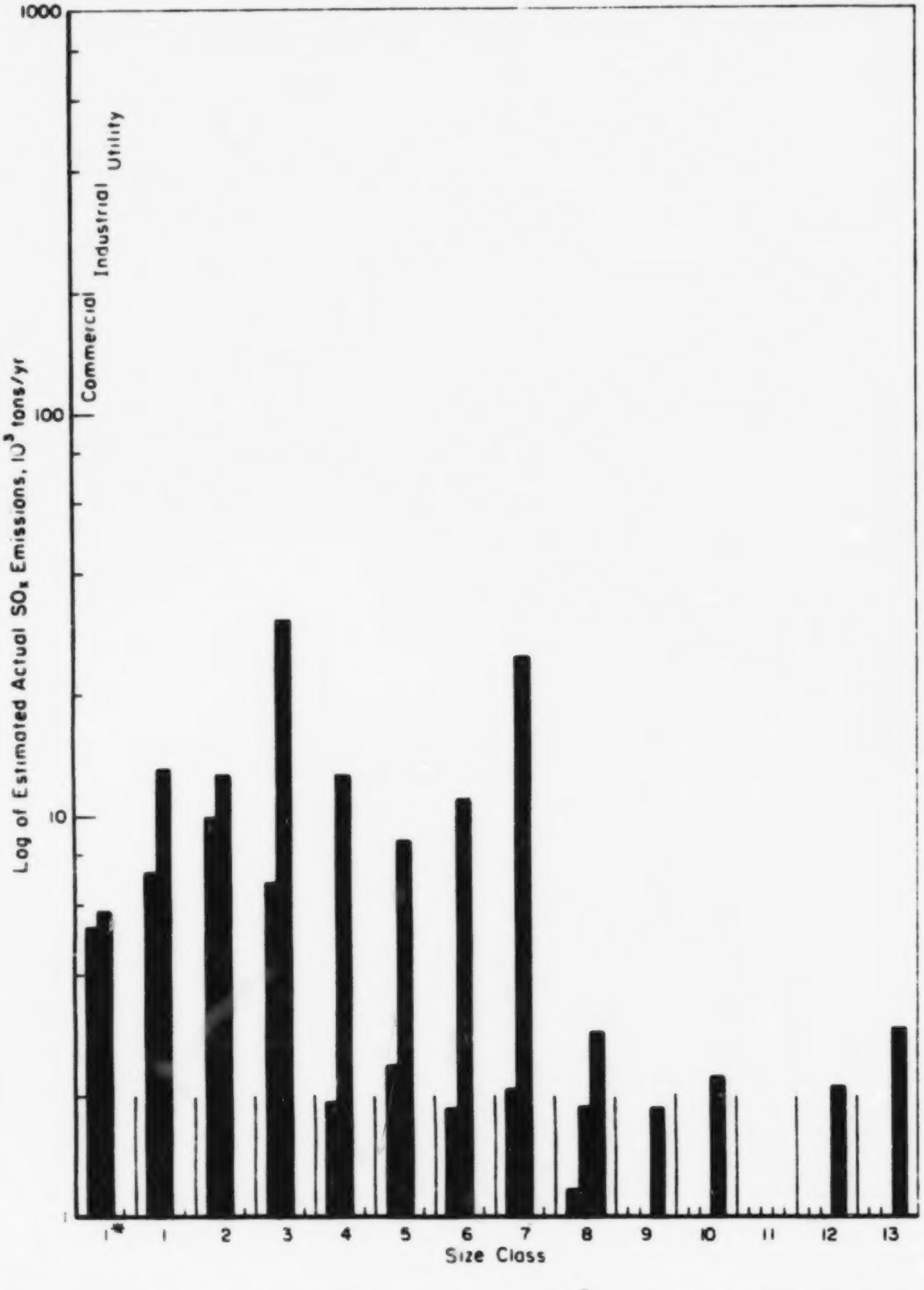


FIGURE 17. ANNUAL HOURLY FUEL USE (10^9 Btu/hr) BY SIZE AND USE CATEGORY FOR LOW SULFUR DISTILLATE OIL FUEL

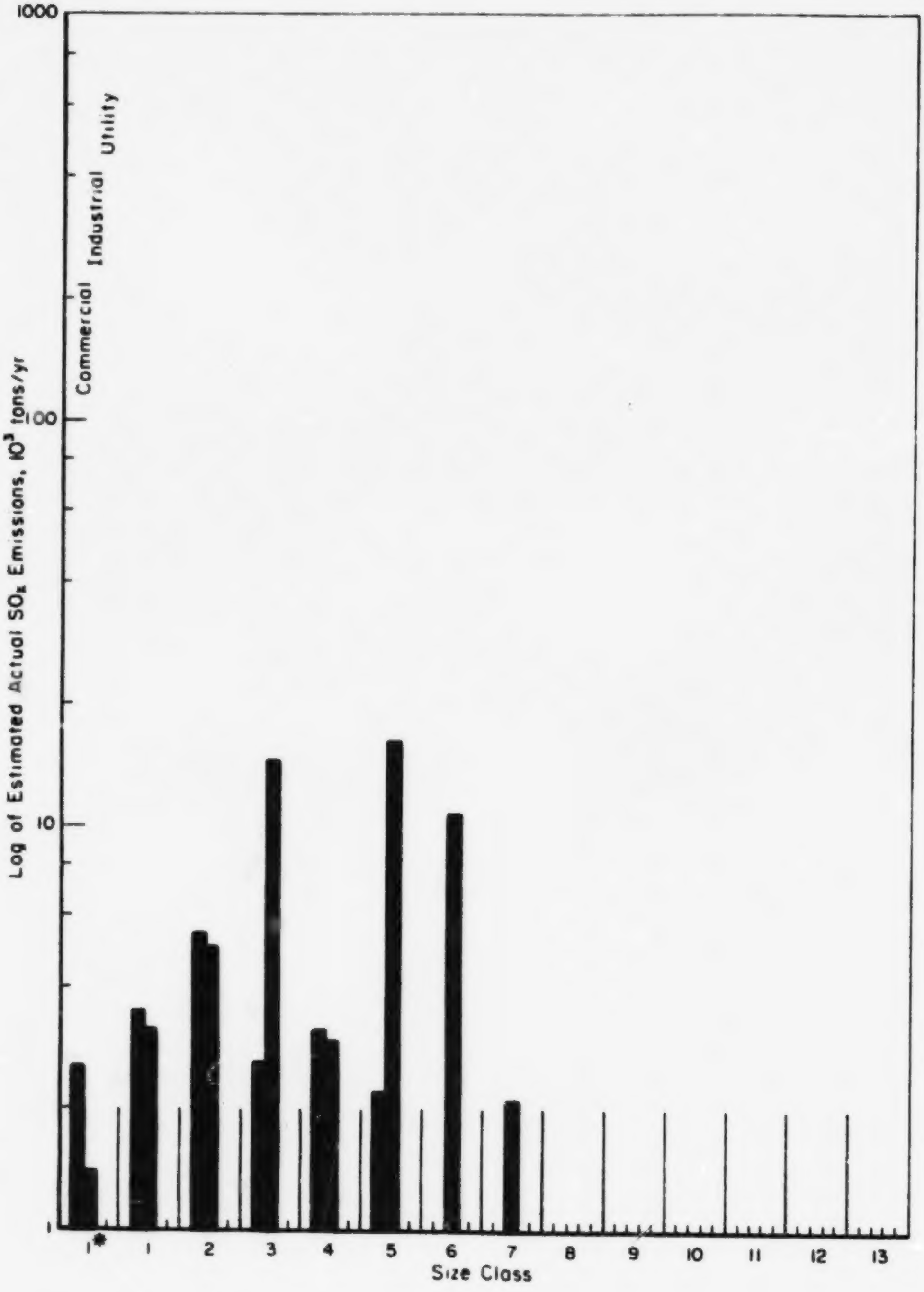


FIGURE 18. ANNUAL HOURLY FUEL USE (10^9 Btu/hr) BY SIZE AND USE CATEGORY FOR HIGH SULFUR DISTILLATE OIL FUEL

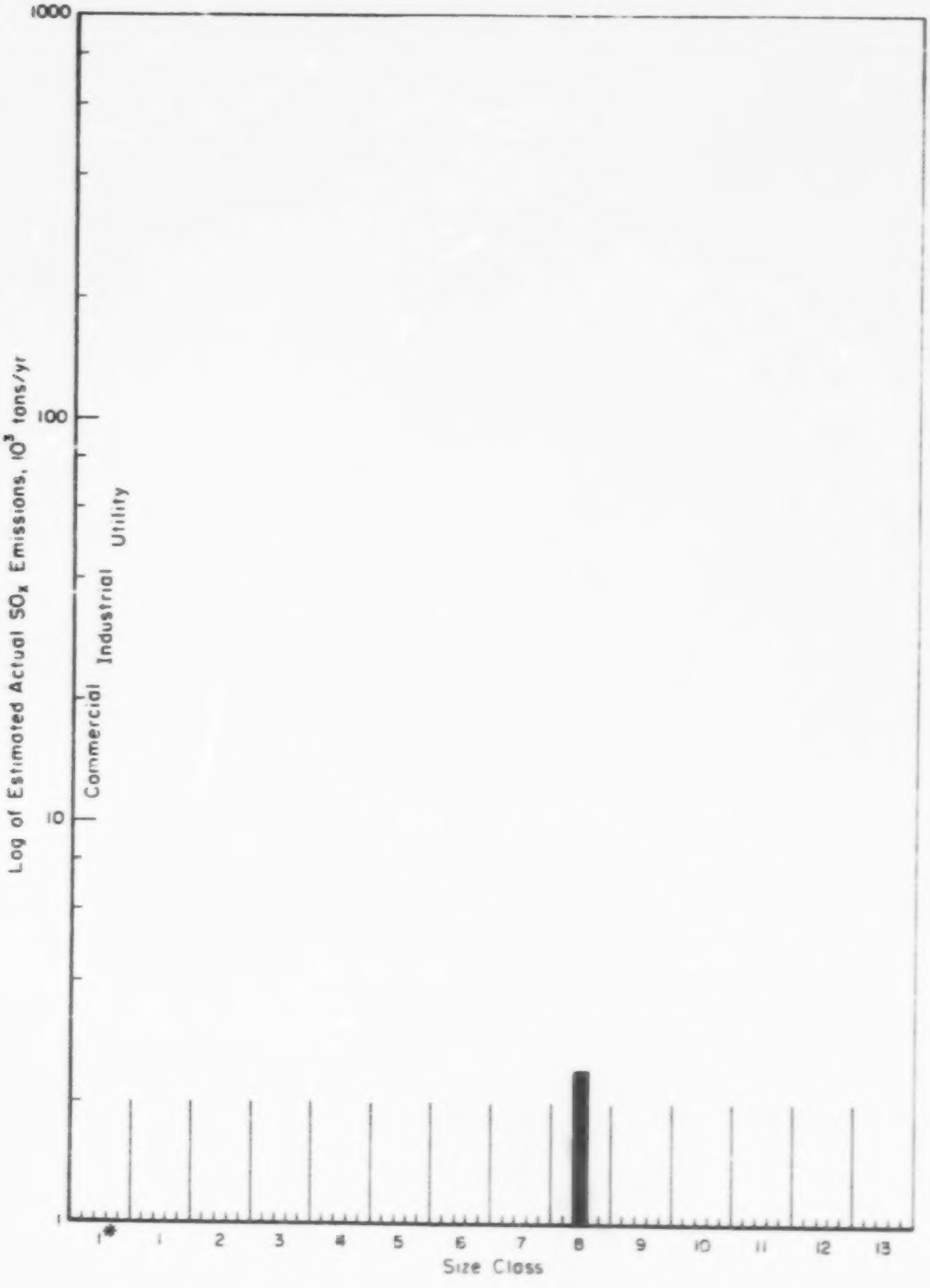


FIGURE 19. ANNUAL HOURLY FUEL USE (10^9 Btu/hr) BY SIZE AND USE CATEGORY FOR HIGH SULFUR NON-NATURAL GAS FUEL

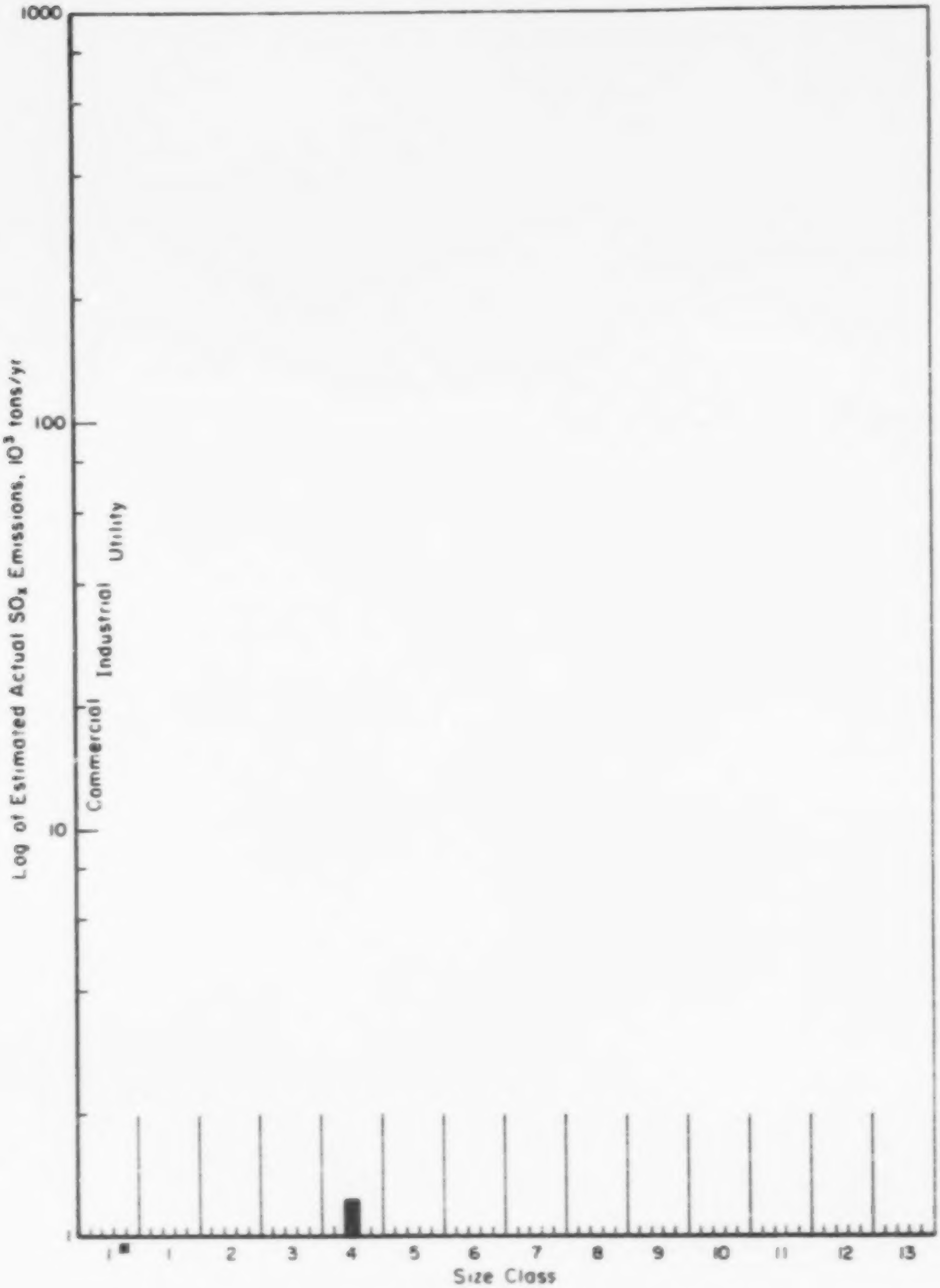


FIGURE 20. ANNUAL HOURLY FUEL USE (10^9 Btu/hr) BY SIZE AND USE CATEGORY FOR LOW SULFUR NATURAL GAS FUEL

PART II

NON-UTILITY COMBUSTION SOURCE CONTROL ALTERNATIVES

CONTROL ALTERNATIVES

The control alternatives considered in this study included

(1) Clean Fuels

- Natural gas
- Low sulfur oil
- Low sulfur coal

(2) Processed Fuels

- Physical cleaning of coal
- Coal gasification
- Coal liquefaction

(3) Combustion Modification

- Fluidized bed combustion of coal

(4) FGD Processes

- Limestone slurry
- Lime scrubbing
- Double alkali
- MgO-central regeneration
- MgO-integrated
- Wellman-Lord

For each of these alternatives the objective was to assess the applicability to small industrial and commercial boilers in terms of the underlying chemistry, operation and maintenance, secondary emissions, raw material requirement, retrofitability, economics, and extent of development.

CLEAN FUELS

Supply Projections

The available supplies of the various clean fuels were estimated for the years 1975, 1980, and 1985. These clean fuel supply projections include the natural clean fuels (natural gas, low sulfur coal, and low sulfur fuel oil) plus desulfurized fuel oil. The quantities used for transportation and for petrochemical feedstocks and other nonfuel uses have been excluded. Thus, these are the supplies available to the residential and commercial, industrial, and utility sectors.

Table 10 shows the clean fuel supply projections. In the upper portion of the table values are given in the usual units and in the lower portion they are given in terms of the fuel heating value. These projections are based on some previous estimates made by Battelle.(9) The following explanation gives the basis of the original estimates and the changes made in this report.

The gaseous fuel supply including natural gas, pipeline imports, and LNG imports was used as provided by Dupree and West.(10) The only petroleum products considered applicable to industrial and commercial boilers were distillate and residual fuel oils. Most distillate fuel oil contains less than 1 percent sulfur by weight. The Minerals Yearbook 1971(11) indicated that distillate fuel oil accounted for 17.5 percent of the total consumption of petroleum products in 1971. This percentage was assumed to hold constant through 1985, and thus the distillate fuel supply was estimated using Dupree and West's(10) projection of total petroleum supply.

According to a study by Hittman Associates, Inc.,(12) the U. S. supply of low sulfur (< 1 weight percent S) residual fuel oil in 1970 was 1.07×10^6 bbl/day. This includes oil from both domestic and foreign sources. A growth rate of 10 percent per year was estimated through 1980, the growth rate then decreasing to 5 percent per year. The high initial growth rate is attributed to the following:

TABLE 10. CLEAN FUEL SUPPLY PROJECTIONS

Type of Fuel	Sulfur Content (weight percent)	Units	Supply Projection for Year		
			1975	1980	1985
Gas		10^{12} scf	23.5	24.8	26.0
Distillate fuel oil	<1.0	10^9 bbl	1.07	1.28	1.54
Residual fuel oil	<1.0	10^9 bbl	0.63	1.01	1.29
Coal	<0.7	10^6 tons	150	170	220
Gas		10^{15} Btu	23.5	24.8	26.0
Distillate fuel oil	<1.0	10^{15} Btu	6.2	7.5	9.0
Residual fuel oil	<1.0	10^{15} Btu	3.8	6.1	7.7
Coal	<0.7	10^{15} Btu	3.6	4.2	5.5
TOTAL			37.1	42.6	48.2

- The fuel demand for the industrial and electrical sectors will depend heavily on low sulfur resid until coal conversion and flue gas desulfurization technologies become commercialized;
- South American oil refineries have shown a willingness to invest in and operate desulfurization plants. Hittman projects a growth rate of 15 percent per year for such facilities through 1980.

The supply projection for low sulfur coal was based on two studies of the distribution of sulfur content of coal - a survey of coal availability by Hoffman, et al.(13) and a Bureau of Mines report on coal shipments.(14) Data based on the latter report are shown in Table 11 . These studies indicate that coal containing less than 1.0 percent sulfur constitutes about 33 percent of the total coal production on a heating value basis or about 39 percent on a weight basis. To adjust these percentages to the desired break point of 0.7 percent sulfur, another fact from the Hoffman report was used, namely, that the recoverable coal reserves containing less than 0.7 percent sulfur are about 67 percent of the reserves containing less than 1.0 percent sulfur. These percentages were used with Dupree and West's projection of total coal supply to estimate the low sulfur coal supply.

Utilization and Applicability

One approach which could be used to reduce the sulfur oxide emissions from an industrial or commercial boiler using high sulfur coal or oil would be to switch to a clean fuel. Natural gas is assumed to be unavailable for additional use in industrial and commercial boilers. Residential uses have the first priority for this fuel, and as the supply situation for natural gas has become very tight, industrial consumption has been cut back. In most areas of the country, gas companies are not accepting any new customers. Thus, only the following fuel switching possibilities are of interest here:

(1) From high sulfur coal to low sulfur coal

(2) From high sulfur coal to low sulfur oil

(3) From high sulfur oil to low sulfur oil.

TABLE 11. SULFUR DISTRIBUTION IN BITUMINOUS AND LIGNITE COAL - BUREAU OF MINES, 1970

Based on Shipments of Coal

Region	Shipments Accounted for (10^6 tons)	Percent of Coal with Sulfur Content (weight percent)				
		<0.5	0.5 - 1.0	1.0 - 1.3	1.3 - 1.8	>1.8
Appalachia	410.6	2.3	45.3	6.4	11.8	34.2
Midwest	143.7	--	4.0	1.0	3.0	92.0
Near West	9.5	1.8	10.4	--	19.1	68.7
Far West	34.2	20.4	70.0	8.7	0.5	0.4
Entire U.S.	598.0	2.8	36.2	5.1	9.2	46.7

Note that a switch from high sulfur oil to low sulfur coal is not included because the alterations required to change a boiler system from a liquid fuel to a solid fuel would be so extensive as to render this change impractical.

Costs

The costs associated with fuel switching are of three types:

- Operating costs due to the difference in price between the two fuels involved
- Boiler modification costs, which include an investment and the associated investment-related annual costs
- Other effects on the boiler operating cost (excluding fuel) due to a change in the state of the fuel.

These costs can vary widely depending on the particular features of the boiler involved, the sources of the fuels involved, and the nature of the fuel purchase agreements. Some generalizations will be made in the following discussion.

Table 12 presents some average cost data for fuels sold to utilities at two time periods, mid-1973 and the end of 1974. Note that in mid-1973 the difference in cost between high sulfur and low sulfur coal was about \$2.30/ton (9¢/10^6 Btu) and the difference between high sulfur and low sulfur residual oil was about \$0.30/bbl (5¢/10^6 Btu). The corresponding values at the end of 1974 were about \$5.70/ton (23¢/10^6 Btu) for coal and \$1.90/bbl (32¢/10^6 Btu) for residual oil.

The present uncertainties in the energy area make projections of fuel costs very difficult and subject to considerable speculation. Particularly important are the questions of whether the price of natural gas will be deregulated and to what extent the U.S. will continue to import foreign oil.

TABLE 12. COST OF FUELS

Type of Fuel	Weight Percent S	Mid-1973 Cost(15)		End-1974 Cost(16)	
		Fuel Units	$\$/10^6$ Btu	Fuel Units	$\$/10^6$ Btu
Natural gas		43¢/10^3 scf	0.43	61¢/10^3 scf	0.61
Distillate fuel oil		$5.70/bbl	0.98	$12.00/bbl	2.06
Residual fuel oil	<1.0	$4.20/bbl	0.70	$12.30/bbl	2.05
Residual fuel oil	>1.0	$3.90/bbl	0.65	$10.40/bbl	1.73
Coal	<0.7	$12.10/ton	0.50	$28.50/ton	1.18
Coal	>0.7	$9.80/ton	0.41	$22.80/ton	0.95

In considering boiler modification costs, note that the first and third fuel switching possibilities listed above involve no modification cost because no change in the state of the fuel is involved. Only the switch from high sulfur coal to low sulfur oil requires boiler modifications. Although no direct, general estimates of this investment are available, an approximation can be made based on some data given by Schreiber, et al.(17) These data are for the investment required to convert boilers from coal to liquid Solvent Refined Coal (SRC). This is a change from a solid to a liquid fuel, although the SRC requires a considerable amount of heating to keep it in a liquid state, thus increasing the investment substantially. Figure 21 shows the investment for the conversion from coal to liquid SRC as a function of boiler capacity. This line represents the consensus of the data given by Schreiber after adjustment to a general U.S. location (using location factors given by Schreiber) and to mid-1973 (using the Chemical Engineering Plant Cost Index). The materials/labor ratio for this investment is about 37/63. The investment for converting from coal to fuel oil will be less than that shown in Figure 21 , and this difference will vary widely depending on the properties of the fuel oil. Heavy residual oils require some heating to maintain flow (although less than SRC) and the investment for these fuels should be not too much lower than the values shown in Figure 21 (probably about 10 percent lower). For distillate fuel oils, which require no heating, the investment would be much lower than the values shown in Figure 21 (probably at least 50 percent lower).

Associated with the investment for boiler modifications is, of course, an investment-related annual cost. If one assumes that there is no maintenance associated with this investment, the annual cost will include only the local taxes and insurance, depreciation, return on rate base, and Federal income tax. For the Utility Financing Method being used here (see Appendix A), these costs amount to annual cost of 14.6 percent of the investment.

There are other effects on the boiler operating cost related to the change of state of the fuel. A survey published by Olmsted(18) showed the following average operating costs excluding fuel for power plants in 1973.

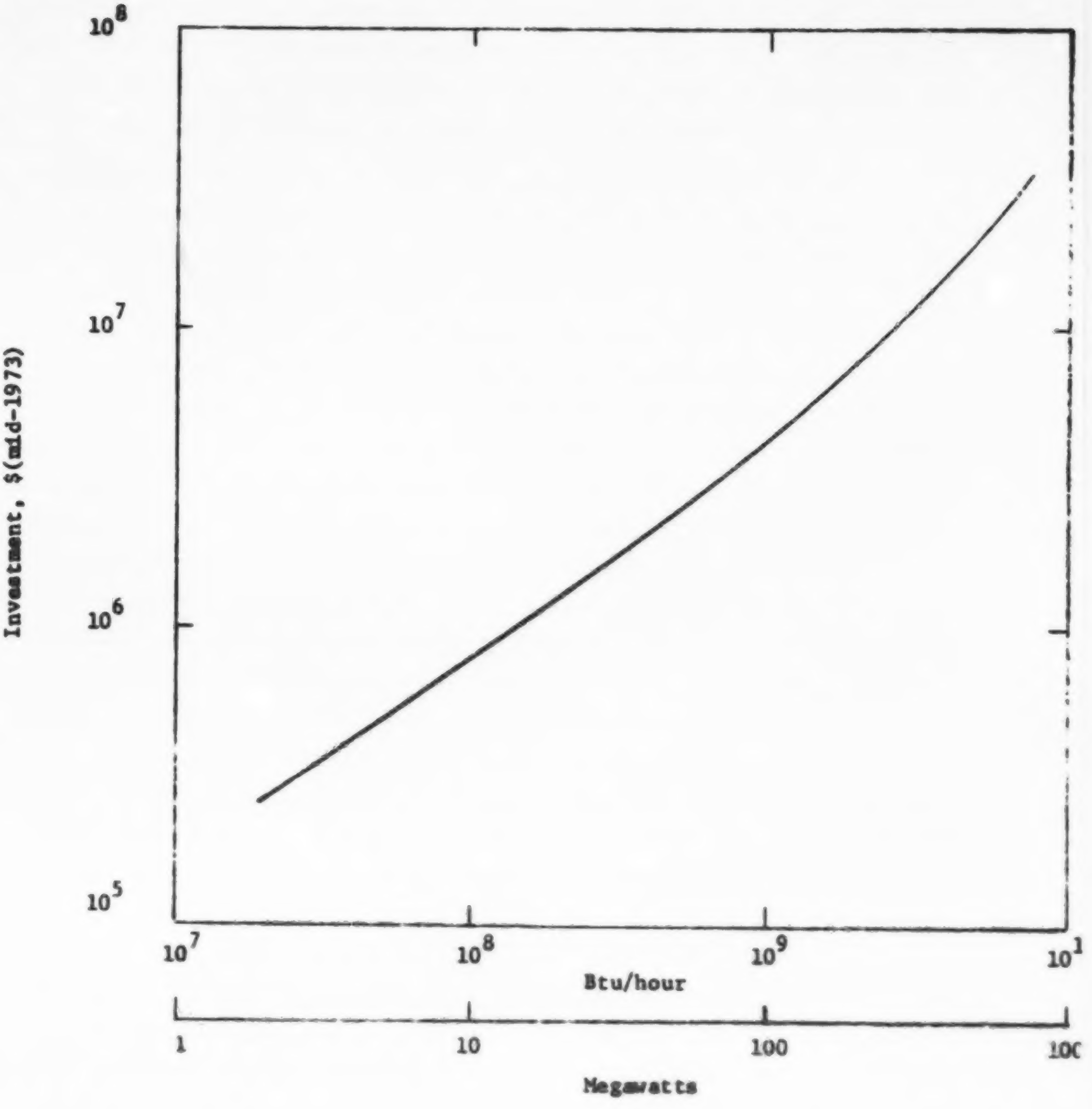


FIGURE 21. INVESTMENT FOR CONVERTING BOILERS FROM COAL TO LIQUID SRC

Fuel Type	Operating Cost Excluding Fuel (mills/kWh)
Gas	0.56
Oil	0.95
Coal	0.75

Note that the operating cost excluding fuel is 0.20 mills/kWh higher for an oil-fired boiler than for a coal-fired boiler. This increase in operating cost must be included when estimating the cost of switching from coal to oil.

PHYSICAL CLEANING OF COAL

The sulfur oxide emission standards for industrial boilers may be met in a variety of ways. The use of low sulfur coal is, of course, a distinct alternative. Assuming an allowable emission factor of 1.2 lb SO_2 per 10^6 Btu, the allowable sulfur content in a coal of heating value of 12,000 Btu/lb will be about 0.7 percent.

In a number of coal-producing regions in the continental United States, some coals are amenable to reduction of their sulfur contents by the so-called "physical" techniques. These techniques utilize differences in physical properties of the coal and the refuse (including minerals of sulfur) such as specific gravity.

Sulfur in Coal

Sulfur is present in coal in several forms. In the organic form it is chemically bonded to the carbon atoms and as such cannot be removed by physical means. In a number of Eastern United States coals, it represents approximately 20 to 85 percent of the total sulfur present.(19) The inorganic form is present mainly as the chemical species pyrite or marcasite (FeS_2) with relatively small amounts of the sulfates of calcium and iron, usually in the range of 0.07 to 0.2 percent. The total sulfur content of coal varies from less than 1.0 to more than 9.0 percent. Also, there is a large variability in the percentage of physically removable pyritic sulfur, not only in coals from different regions, but also in coals obtained from the same mine. The average organic sulfur content of a number of Eastern U.S. coals has been reported at 51.2 percent(19) of the total sulfur.

Cleanability of U.S. Coals

The coal resources of the United States are found in three belts or regions, namely, the Appalachians, the Eastern Interior, and the Western (Rocky Mountain). It has been found that the bituminous coals from the Northern Appalachian Region show a greater propensity for sulfur reduction by physical means. The 0 to 0.7 percent sulfur category with total reserves (as of 1968) of 43.8 million tons may "grow" to 9,930 million tons by crushing higher sulfur coals to a 3/8-inch top size and washing such that a 90 percent yield is obtained. In the remaining coal-producing regions, physical cleaning does not materially change the sulfur content. The coals in the Western Interior are "characteristically high in (organic) sulfur" while the coals in the Northern and Southern Rocky Mountain regions are inherently of low sulfur content.(20)

A recent analysis of the sulfur contents of 200 samples from the Northern Appalachian region showed an average total sulfur content of 3.0 percent. Crushing to a top size of 3/8 inch followed by washing reduced this average to 1.6 percent.(21) A more recent analysis(22) has shown that in the Northern Appalachian region (Maryland, Ohio, Pennsylvania, and Northern West Virginia) less than 5 percent of the samples would comply with the EPA standard of 1.2 lb/10^6 Btu. Only 35 percent of the samples would comply after crushing to 14 mesh top size and washing to a point where 50 percent of the Btu value is recovered. The corresponding figures for Midwest Region coal (Illinois, Indiana, and Western Kentucky) are 1 percent and 5 percent, respectively.

Technology of Coal Cleaning

The principal coal preparation processes used today are oriented toward product standardization and ash and sulfur (pyritic) reduction.

In a preparation plant the raw coal is typically subjected to (1) size reduction and screening, (2) separation of coal from ash and pyrite in a device utilizing differences in specific gravity or surface properties between them, and (3) dewatering the product coal and refuse.

Size reduction is accomplished in rotary or roll crushers. The extent of preliminary and subsequent (farther along in processing) size reductions depends on the type of coal processed and the desired product characteristics, mainly ash and pyrite and Btu content. It is a well known fact that more of the impurities are liberated as size of the coal is reduced. An economic optimum is usually sought as the costs of preparation rise exponentially with the percentage of "fines" to be treated.

Screening (either wet or dry) is practiced to separate the various size fractions resulting from the crushing operations. Separations based on differences in specific gravity are usually reserved for fractions Mesh No. 35 or over.

Separation of pyrite and ash from the raw coal may be accomplished in a variety of devices or processes. Jigging currently handles the largest percentage of coal cleaned by physical means.

Dewatering operations are carried out in equipment such as fixed and vibrating screens, thickeners, centrifuges, vacuum fillers, and thermal dryers.

Environmental Impacts and Control

Air Pollutants. Particulate matter originating from the operation of thermal dryers, and coal crushers is the most serious pollutant from coal preparation plants. With increased percentage of fine coal (Mesh No. 35 and under) being processed, the uncontrolled emissions of particulate matter per ton of total product have increased because the finer sizes are usually thermally dried. Uncontrolled emissions of particulate matter may be in the range of 15 to 25 pounds per ton of thermally dried coal, depending upon the type of dryer used.(23) The thermally dried coal may constitute two-thirds the total plant output. The uncontrolled emissions of particulate matter may thus be as high as 10 to 17 pounds per ton of clean coal. Federal regulations(24) limit the emissions from thermal dryers to 0.031 grains per scf. For a typical dryer, the gaseous discharge is of the order of 37,000 scf per ton of coal dried.(24) This translates into an allowable emission level of 0.16 pound of particulate matter per ton of coal dried.

Other pollutants of concern are SO_x, NO_x, hydrocarbons, and carbon monoxide resulting from the combustion of product coal to effect drying by direct contact of wet coal and the hot gas. These emissions are variable and are not regulated. The variability results mainly from the variability in sulfur content of product coal and efficiency of the combustion device used. If a heat requirement of 230 Btu per pound coal dried is assumed, then the emissions per ton product may be calculated using published emission factors on industrial boilers burning bituminous coal.(24) These emission factors in pounds per ton of coal burned are: 1 for carbon monoxide, 15 for nitrogen oxides (as NO_2), and 1 for hydrocarbons. Assuming a sulfur content of 1.5 percent, 12,000 Btu per pound for product coal, and 67 percent of plant output thermally dried, then the emissions per ton of coal cleaned may be of the order of 0.8 lb SO_2, 0.2 lb NO_x (as NO_2), 0.013 lb carbon monoxide, and 0.013 lb hydrocarbons.

Water Pollution. Large amounts of water circulate in a coal washing plant. Although modern coal cleaning plants are designed to operate on a closed circuit,(19) make-up is needed for water losses in the refuse, the clean coal, and in the thermal dryer. This may vary between 15 and 40 gallons per ton of clean coal,(26) depending on age and type of plant. The characteristics of water circulating in a coal cleaning plant are similar to those of acid mine drainage water.

Solid Waste. Solid wastes from a coal preparation plant are mainly the pyrite and ash-forming refuse together with the coal value lost in preparation. These may amount to 300 lb per ton of clean coal(26) and an associated surface moisture of 70 lb per ton. These quantities vary with such factors as ash content of coal, product specifications, and type of washing operation employed. Appropriate methods of disposal are discussed in Reference 19.

Applicability to NUC Sources

The physically cleaned coal is a solid fuel and thus the application to industrial boilers is limited to the boilers that burn coal. Although the cleanability of U. S. coals is somehow limited due to the physical and chemical characteristics of coal, if coal is accessible to cleaning, this alternative is very favorable economically to coal-fired industrial boilers which in general are small in size.

Capital and Annualized Costs

The capital costs presented in Figure 22 are mid-1974 for a modern plant including appropriate emission and effluent control systems and refuse handling systems.(26) The breakdowns of total capital requirement and annual cost are shown in Table 13. The estimations were based on the Utility Financing Method as modified by the Panhandle Eastern Pipeline Company(27) (see Appendix A) unless specified in the table and the footnotes. The product coal cost was estimated at $13.58/ton and $13.25/ton for a 500 tph and 1500 tph capacity plant, respectively.

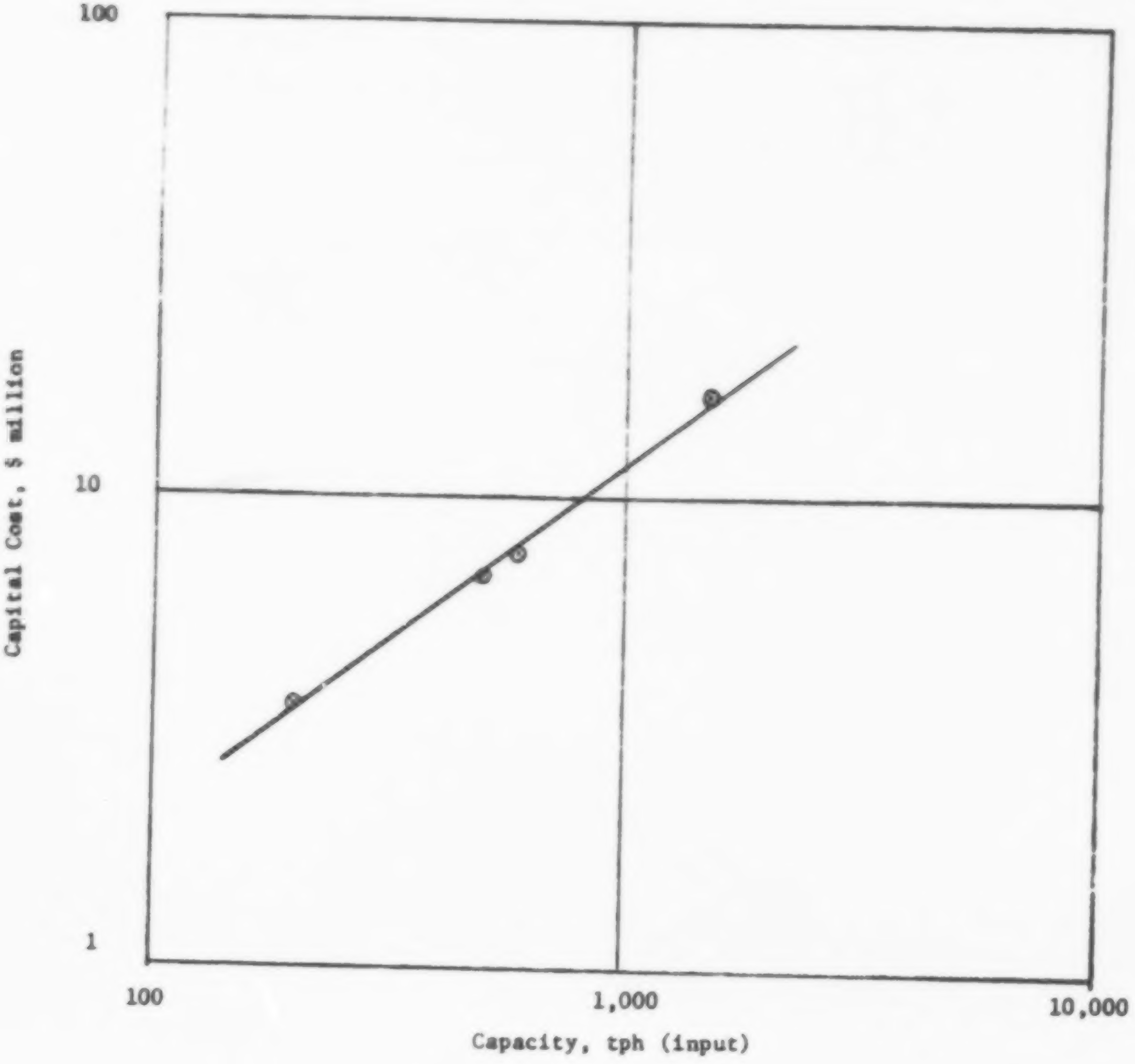


FIGURE 22. CAPITAL COSTS OF COAL WASHING PLANTS MID-1974

TABLE 13. COSTS FOR PHYSICAL COAL CLEANING

Mid-1973, 4000 hours/yr Operation

Item	Quantity	
Coal Input Rate (tons/hr)	500	1,500
Clean Coal Product Rate (tons/hr)	429	1,287
Capital Requirement (1000$)		
Total bare cost(a)	6,025	13,583
Engineering and design	301	679
Contractor fees	603	1,358
Subtotal Plant Investment	6,929	15,620
Project contingency	693	1,562
Total Plant Investment	7,622	17,182
Interest during construction	1,286	2,899
Startup	4,227	12,515
Working capital	4,227	12,515
Total Capital Requirement	17,362	45,111
Annual Costs (1000$/yr)		
Direct operating labor(b)	120	200
Maintenance (5 percent/yr)(c)	381	859
Supervision	43	87
Admin. and general overhead	199	399
Local taxes and insurance	206	464
Electricity (1¢/kwhr)(d)	119	358
Lime (1.4¢/lb)(e)	10	31
Magnetite (2¢/lb)(e)	13	40
Frothing agent (22¢/lb)(e)	14	42
Flocculant (3¢/lb)(e)	31	92
Water (2¢/1000 gal)(e)	1	2
Annual Cost ex. Coal	1,137	2,574
Coal cost $10/ton)	20,000	60,000
Annual Operating Cost	21,137	62,574
Depreciation	657	1,630
Return on rate base	1,133	3,025
Federal income tax	374	998
Total Annual Cost	23,301	68,227
Product Cost ($/ton product)	13.58	13.25
Product Cost ($/$10^{6}$ Btu output)(f)	0.543	0.530
Heat Output Rate (10^{12} Btu/yr)	42.9	128.7

Footnotes to Table 13

(a) 1973 Total Bare Cost ($) = 60,630 (tons/hr input)$^{0.74}$.

(b) Operating labor = 6 men/shift for 500 tons/hr, 10 men/shift for 1500 tons/hr.

(c) Maintenance materials/labor ratio = 56/44.

(d) Power consumption = 4000 H.P. for 500 tons/hr.

(e) Materials consumptions:

Lime	2.6 lb/T dry solids
Magnetite	0.5 lb/T coal treated by dense medium
Frothing agent	0.17 lb/T coal treated by froth flotation
Flocculant	3.6 lb/T dry solids
Water	17 gal/T clean coal

From material balance flow sheet:

T clean coal/T coal input = 0.858
T dry solids/T clean coal = 0.165
T coal by dense medium/T clean coal = 0.783
T coal by froth flotation/T clean coal = 0.217
T coal dried/T clean coal = 0.662

Heat required for dryer = 230 Btu/lb coal dried.

(f) Based on clean coal heating value of 12,500 Btu/lb.

COAL GASIFICATION

Gasification of coal is intended to produce gaseous products for consumption as fuels and for industrial uses such as chemical synthesis and reducing gas in iron and steel production. The technology has been practiced in a commercial scale primarily in Europe for many years to manufacture town gas, fuel gas, and synthesis gas. Gasification of coal involves the reaction of coal with air, oxygen, steam, CO_2, or mixtures of these gases. A low-Btu gas is obtained if an air-steam mixture is used directly to gasify the coal and contains nitrogen as a major component. Intermediate-Btu gas, which contains a minor amount of nitrogen, is obtained when an oxygen-steam mixture is used. High-Btu gas, which is similar to natural gas and contains over 90 percent methane, is obtained by further processing of intermediate-Btu gas. Low-Btu gas is suitable for use as an energy source near its point of generation, but it is not economically favorable for long distance transportation. Intermediate-Btu gas can be used either as an energy source or as a synthesis gas for the production of chemicals. For analyses in this study, the low- and intermediate-Btu gases were categorized as low-Btu gas to distinguish them from methanated high-Btu gas.

Gasification Processes

Low-Btu Gas. In general, the process consists of coal preparation, gasification, gas cleaning, desulfurization, and compression. Mined coal is crushed, screened, and then conveyed to the storage bunkers atop coal lock hoppers. As the coal is fed to gasifiers, it reacts with externally supplied oxygen and steam, and as the result hydrogen, carbon monoxide, carbon dioxide, and methane are produced. The steam is the source of hydrogen and the heat resulting from combustion of coal supplies the heat required for gasification. In addition to coal gasification, coal devolatilization or carbonization takes place in the reactor. Gaseous products of devolatilization are rich in methane and hydrogen and contain tars and oils.

The crude gas is quenched and scrubbed by a wash cooler and then desulfurized before utilization.

Various gasification processes can be used to produce low-Btu fuel gas. Presently, four gasification systems can be considered commercially viable in the sense that their technologies are commercially proven and their systems are available through commercial dealers.(28) These four systems include

(1) Lurgi

(2) Koppers-Totzek

(3) Winkler

(4) Wellman-Galusha.

None of these processes are in widespread use in the United States at the present time mainly due to the past abundance of inexpensive natural gas and fuel oils which today are becoming increasingly scarce.

High-Btu Gas. Basically, a high-Btu gasification process is identical to the low-Btu gasification process except for shift reaction and methanation processes. The scrubbed crude gas is introduced to the shift reactor. About one-half of the total gas is subjected to shift conversion. The resulting hot gas is then cooled to facilitate subsequent purification by an acid gas removal process. The product gas from acid gas removal process is fed to a fixed-bed methanation reactor. The methanated gas is compressed and dehydrated for pipeline gas.

Various coal gasification processes have been developed for manufacturing high-Btu synthetic natural gas (SNG). Currently, none of these have been constructed for commercial operation. The processes in an advanced stage of development include Lurgi, Synthane, Hygas, and CO_2-Acceptor. The Lurgi process is commercially proven for high-Btu SNG. The full-scale evaluation of the methanation process in conjunction with the Lurgi process at Westfield, Scotland, indicates that the methanation is now commercially feasible.

Fuel Gas Desulfurization and Sulfur Recovery

Low-Btu Gas. The hot raw gas from coal gasification contains many impurities. The primary impurity of environmental concern is sulfur in the form of hydrogen sulfide (H_2S) with very small amounts of carbonyl sulfide (COS) and carbon disulfide (CS_2). In general, the coal gasification process has an advantage over direct combustion of coal in the control of sulfurous compounds since the reduced form of hydrogen sulfide is more easily removed than the oxidized form of sulfur dioxide in flue gas. In addition, removal processes are commercially available for hydrogen sulfide while sulfur dioxide diluted in large volumes of flue gas presently cannot be effectively controlled by commercially-proven processes.

The concentration of hydrogen sulfide in fuel gas depends on the sulfur content in the coal, the heating value of gas, the heating value of coal, and gasification efficiency. Assuming that a coal with a heating value of 12,000 Btu/lb on a moisture- and ash-free basis is gasified with a gasification efficiency of 70 percent, and that at least 90 percent of the sulfur in the coal is converted to hydrogen sulfide, the concentration of H_2S in fuel gases with different heating values can be plotted against sulfur content in coal as shown in Figure 23. Figure 24 shows the emission of SO_2 in pounds per 10^6 Btu fuel gas heat input as a function of H_2S concentration and fuel gas heating value. It can be seen from Figure 24 that a concentration of about 700 ppm and 2100 ppm would be the allowable concentration of H_2S for the fuel gases with heating values of 100 Btu/scf and 300 Btu/scf, respectively, to meet the Federal standard for combustion sources of 1.2 lb $SO_2/10^6$ Btu heat input.

Control of sulfurous emissions can be considered under three broad steps(29):

(1) Desulfurization of fuel gas
(2) Sulfur recovery
(3) Tail gas treatment.

Desulfurization of fuel gas, is usually accomplished by absorption into a liquid phase using suitable gas-liquid contacting equipment.(29) Absorption processes can be divided into three broad categories depending

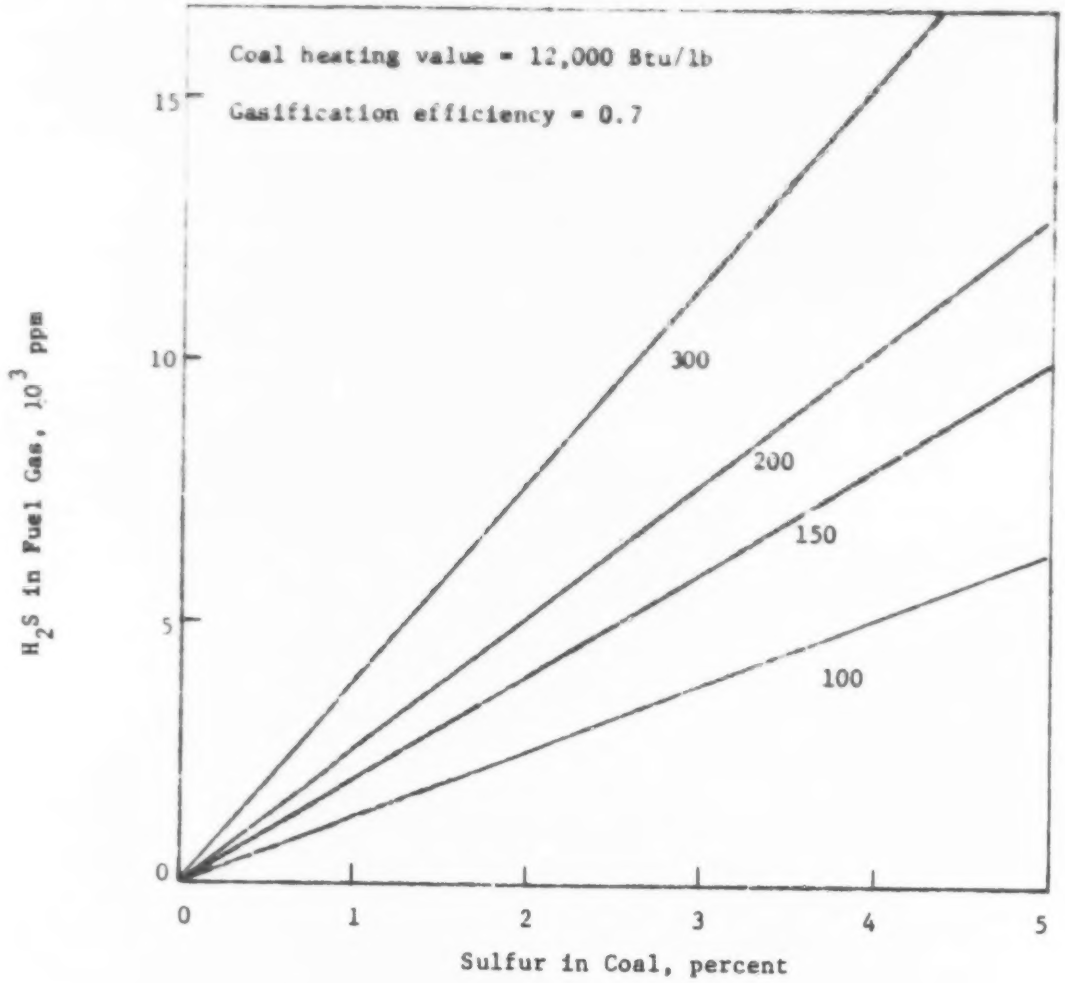


FIGURE 23. H_2S IN FUEL GAS AS A FUNCTION OF SULFUR IN COAL (Parameter: Gas heating value, Btu/scf)

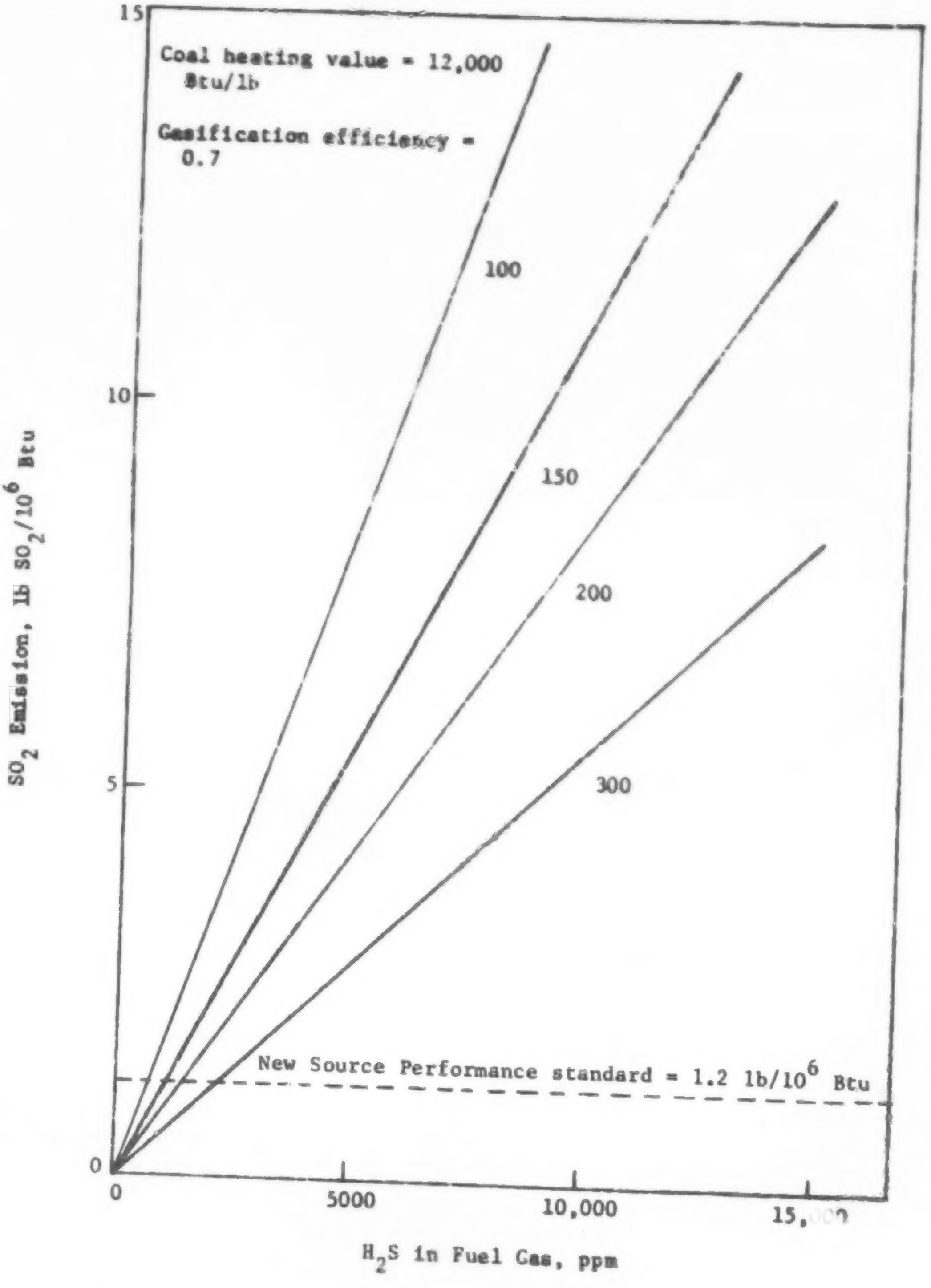


FIGURE 24. SO_2 EMISSION VERSUS H_2S CONCENTRATION IN FUEL GAS (Parameter: Gas heating value, Btu/scf)

upon the type of absorbent used such as physical solvent process, amine process, and alkaline salt solution process. There are direct conversion processes, such as the Stretford and Giammarco-Vetrocoke processes, which are capable of absorption and oxidation of H_2S to produce sulfur directly.

Commercially important processes which recover elemental sulfur from sulfur-bearing gases such as fuel gas and regenerated acid gas include the Claus process, the Stretford process, and the Giammarco-Vetrocoke.(29) All three processes are designed normally to convert hydrogen sulfide in the feed gas to sulfur.

The tail gas from the Claus plant contains unconverted H_2S and SO_2 and lesser quantities of other sulfur constituents, such as COS, CS_2, and elemental sulfur vapor and particles. The tail gas from typical Claus plant operations contains about 1 to 2 percent total sulfur. The normal practice in the past has been to discharge the tail gas directly to atmosphere after passing it through an incinerator to convert sulfur compounds into sulfur dioxide. During the past several years, a number of treatment processes have appeared for removing the residual sulfur compounds from the tail gas. Tail gas treatment processes which have received commercial acceptance are Beavon, Cleanair, IFP-1, SCOT, and Wellman-Lord.(29)

High-Btu Gas. Like a low-Btu coal gasification process, control of sulfurous emissions can be considered under the three steps described above except for the desulfurization of fuel gas. The desulfurization process for high-Btu gasification not only removes sulfurous components but also carbon dioxide in the fuel gas. Removal of carbon dioxide is necessary to avoid undesired corrosion in the pipelines.

Applicability to NUC Sources

Low-Btu Gas. The applicability of low-Btu gas to existing coal-, oil-, or gas-fired boilers will technically depend on boiler configuration and operation. For a given heat input rate, the volumetric flow rate of low-Btu gas is high compared with that of high-Btu gas such as natural gas and so is the combustion flue gas flow rate. These conditions would create the following problems:

(1) The size of piping or ducts to transport the fuel will increase as the heating value decreases.

(2) The burners to handle very low-Btu gas will be different from the existing burners. For oil or high-Btu gas, a small portion of the throat area is required for fuel and most of the burner throat area is for the combustion air. For a low-Btu gas with high volumetric flow rate, it is necessary to increase the burner throat size or the number of burners.

(3) The heat absorption pattern may be different. In general, heat is transferred from the products of combustion to the boiler by radiation and convection. When a low-Btu gas is applied, the rate of heat transfer depends on the temperature and the mass flow of the flue gas over the heating surface. Any change in flue gas temperature or flue gas quantity in a boiler affects the heat transfer balance.

According to the result of the industrial and commercial boiler data analysis conducted in this study, industrial boiler subgroups of environmental concern include small and large industrial boilers burning high-sulfur coal and resid. Therefore, the application pattern of low-Btu gas to existing boilers may be considered only for the coal- and oil-fired boilers; however, it can also be conceived that the application may be necessary to natural gas-fired boilers due to the shortage of the supply.

The application of low-Btu gas to a coal-fired boiler would be more adaptable than to other type boilers. This is mainly because a coal-fired boiler has a large combustion chamber. Although existing stoker coal boilers were designed for maximum utilization of radiation, the conversion to low-Btu gas boiler would not hamper the heat absorption rate because of the increased convective heat transfer. Reduced heat absorption, however, may result for a low-Btu gas with a heating value less than 200 Btu/scf. The pressure loss due to the increased quantity of flue gas may exceed the design condition for gas with a heating value of less than 200 Btu/scf.

The application of low-Btu gas to oil- and gas-fired boilers may be difficult for a gas with a heating value less than 300 Btu/scf. This is mainly because the firing chamber is small and, thus, may not be able to handle a larger volume of combustion gas. Less heat would be absorbed in the furnace and more heat in the superheater and reheater than the unit was originally designed for because of the low temperature of the combustion gas and the increase in loss of sensible heat. The pressure loss would be very high when burning low-Btu gas since the loss is proportional to the square of the flue gas quantity. Some extensive structural alterations may be required to minimize the pressure loss.

The economic applicability of low-Btu gas to existing boilers will depend on load factor, heat loss, and retrofit difficulty. The load factor of industrial boilers falls between 0.35 and 0.50 and, thus, will play an important role in evaluating the economic feasibility. The heat losses from a low-Btu gas boiler will be higher than those for existing boilers due to the increased quantity of flue gas with considerable sensible heat. The retrofit difficulty may be significant for existing boilers with little extra space around the boiler area.

High-Btu Gas. The application of high-Btu SNG to existing coal-, oil-, and gas-fired boilers will be similar to that of natural gas to the boiler systems. The application to natural gas boilers should not present any problem. The application to oil-fired boilers may need a minor rebalancing of heat absorption in the furnace. The application to coal-fired boilers may need a moderate adjustment of the heat absorption system, particularly for the stoker boilers, because the system was originally designed to maximize the radiation heat transfer due to high firing temperature. In general, the conversion of an oil- or coal-fired boiler to a high-Btu gas boiler is not difficult both technically and economically, and, thus, the applicability of this alternative is deemed high.

Model Plant Calculation

Low-Btu Gas. In determining the emission factors and capital and annualized operating costs of low-Btu gasification, a conceptual design study for a plant producing low-Btu gas for five 250,000 lb steam/hr industrial boilers was carried out in this study. The results are listed in Table 14. The sulfur dioxide emissions from the combustion of the fuel gas was estimated at 0.5 lb/10^6 Btu(30) based on the assumption that the fuel gas is desulfurized by a MDEA absorption process with a removal efficiency of 93.5 percent. The sulfur dioxide emission from a combined system of a Claus unit and a Beavon tail gas treatment system was assumed to be about 0.02 lb/10^6 Btu.

High-Btu Gas. To determine the manufacturing cost of high-Btu SNG, a conceptual design study was carried out for a typical SNG gasification process. The Hygas process was selected mainly because the process can use both caking and noncaking coals. The Hygas process under consideration in this study is based on the design by IGT(32) to produce 265 x 10^6 scf/day of pipeline gas using a Pittsburgh seam coal. Since no updated economic data are available in the open literature, the data presented in this study (see Table 15) were based largely on the IGT design. The data, however, were corrected with respect to coal flow rate, sulfur content, and base year. A scaling factor of 0.8 was assumed for sulfur content of coal. The total capital requirement was estimated at $401.8 million and net annual operating cost at $82.8 million. The average gas cost over the life time of the plant of 20 years was estimated to be $1.60/$10^6$ Btu.

TABLE 14. ESTIMATED COST OF THE KOPPERS-TOTZEK LOW-BTU GAS

Gasification Process: Koppers-Totzek
Plant Size: 118 x 10^6 scf/day
Plant Load Factor: 0.9
Gasification Efficiency: 0.7
Gas Heating Value: 300 Btu/scf

Coal Feed Rate: 2,107 tons/day
Type of Coal: Eastern Bituminous
Sulfur Content of Coal: 3%
Ash Content of Coal: 14%
Coal Heating Value: 12,000 Btu/lb

Item	Cost, $\$10^6$
Capital Requirement	
Bare cost[a]	19.7
Engineering and design	1.0
Contractor's overhead and profit	2.0
Subtotal Plant Investment	22.7
Project Contingency	3.4
Total Plant Investment	26.1
Interest during construction	4.4
Startup cost	1.9
Working capital	1.9
Total Capital Requirement	34.3
Annual Operating Cost	
Labor[b]	0.79
Administrative and general overhead	0.49
Materials and utilities[c]	0.87
Fuel Cost[d]	6.92
Local taxes and insurance	0.70
Gross Operating Cost	9.75
Credit[e]	0.19
Net Operating Cost	9.56

TABLE 14. (Continued)

Item	Cost, $\$10^6$
Average Gas Cost	
Return on rate base	1.90
Federal income tax	0.63
Depreciation	1.62
Net operating cost	9.56
Average Annual Cost	13.71
Average Gas Cost, $\$/10^6$ Btu	1.18

(a) The value was obtained from Reference (31). It was corrected with respect to base year by using the CE plant cost index. This cost includes gasification, acid gas removal and sulfur recovery, oxygen plant, and pollution control equipment. This does not include utilities, off-site facilities, and land.

(b) The value was assumed.

(c) This includes maintenance and operating supplies and direct material and utility cost excluding cost for coal.

(d) The cost of coal was assumed at $10/ton.

(e) This includes credits for by-product sulfur at $10/long ton and for reduced operating cost of boiler system.

TABLE 15. ESTIMATED COST OF THE HYGAS SNG

Gasification Process: Hygas	Coal: Eastern coal
Gasification Plant Size: 260 x 10^6 scf/day	Coal Feed Rate: 17,517 tons/day
Plant Load Factor: 90 percent	Sulfur Content of Coal: 3 percent
Gas Heating Value: 963 Btu/scf	Ash Content of Coal: 14 percent
Gasification Efficiency: 60 percent	Heating Value of Coal: 12,000 Btu/lb

Item	Cost, $\$10^6$
Capital Requirement	
Total bare cost(a)	248.6
Engineering and design cost	Included in total bare cost
Contractor's overhead and profit	24.9
Subtotal Plant Investment	273.5
Project contingency	41.0
Total Plant Investment	314.5
Interest during construction	53.1
Startup cost	17.1
Working capital	17.1
Total Capital Requirement	401.8
Annual Operating Cost	
Labor(b)	7.9
Administrative and general overhead	4.7
Materials and utilities(c)	7.0
Fuel cost(d)	57.5
Local taxes and insurance	8.5
Gross Operating Cost	85.6
Credits(e)	2.8
Net Operating Cost	82.8

TABLE 15. (Continued)

Item	Cost, $\$10^6$
Average Gas Cost	
Return on rate base	22.0
Federal income tax	7.3
Depreciation	19.2
Net operating cost	82.8
Average Annual Cost	131.3
Average Gas Cost, $\$/10^6$ Btu	1.60

(a) This includes costs for coal preparation, gasification, shift conversion, acid gas removal and sulfur recovery, oxygen plant, methanation, pollution control, utilities and off sites, and land. The value was obtained from References (32) and (33). It was corrected with respect to coal flow rate and base year by using a scaling factor of 0.9 and the CE plant cost index, respectively, where needed.

(b) This includes direct operating labor, maintenance labor, and supervision. A direct operating labor of 52 men/shift was used as suggested in References (32) and (33).

(c) This was estimated from References (32) and (34). It was corrected with respect to coal consumption rate and base year. This does not include the cost for coal.

(d) The cost of coal was assumed at $10/ton.

(e) This includes the following credits:

Elemental sulfur at $10/long ton	$\$1.5 \times 10^6$
Ammonia at $25/ton	$\$1.1 \times 10^6$
Phenol at $0.02/lb	$\$0.2 \times 10^6$
	$\$2.8 \times 10^6$

COAL LIQUEFACTION

The increased demand for clean fossil fuels has stimulated development of methods for converting the nation's abundant coal resources into a low-sulfur, low-ash fuel. One technique, coal liquefaction, utilizes solvents, heat, and high pressure to "liquefy" the coal to produce an ashless, low-mineral, low-sulfur, high-Btu fuel.

Process Description

The current U.S. coal liquefaction or extraction processes may be classified either as a solvent refining or hydrogenation process. The solvent refined coal programs of the Office of Coal Research and Southern Research Institute, and the dissolved coal variation of the Consolidation Coal Company synthetic fuel process are in the first group; the H-coal, USBM, Gulf, and other catalytic processes fall into the second group.

The H-Coal process involves simultaneous catalytic hydrogenation and dissolution of the coal in a specially designed ebullated reactor. The reactor product slurry is transferred to a flash drum to separate the lighter hydrocarbons from the slurry. The slurry is then passed through hydroclones to separate the recycle solvent. The underflow stream is filtered to remove the minerals and undissolved carbonaceous matter, leaving a liquid stream which may be distilled to separate the naphtha from fuel oil. Hydrocarbon Research states that the ashless liquid product contains about 0.2 percent sulfur and has a heating value of about 18,000 Btu/lb.(35) About 18,600 scf of hydrogen is consumed per ton of coal in processing Illinois No. 6 coal with 5.0 percent sulfur and 9.9 percent ash content. About 2.7 barrels of synthetic crude distillate are produced per ton of coal processed.

The Synthoil process features a packed-bed reactor operating at 840°F and 2000 to 4000 psig in which coal dissolution and catalytic hydrogenation occur simultaneously. The effluent gases are separated from the extract in high-pressure receivers. After pressure let-down, the extract oil is either centrifuged or filtered to remove mineral and undissolved organic matter. The product oil is of reasonably low viscosity and flows freely at room temperature. Hydrogen consumption is maintained relatively low, about 9000 scf/ton of coal, as only enough is added to remove the sulfur. Gas production is also minimized. The major difficulty with the process is the high pressure drop with attendant high pumping requirements. Experimental results indicate that the Synthoil Process is capable of producing three barrels of synthetic oil with a 0.19 percent sulfur content from one ton of 4.6 percent sulfur coal.(36) The liquid has a heating value of 15,000 Btu/lb.

The Gulf process utilizes a fixed-bed reactor specifically designed to minimize catalyst plugging to liquefy and catalytically hydrogenate the coal. The reactor product passes to a gas-liquid separator where hydrogen is recovered for recycle. The liquid product goes to solids separation, normally hydroclones, where the slurry overflow is recycled and the high solids underflow is sent to a solids removal process such as filtration of vacuum distillation. Gulf Research(37) states that their catalytic process will generate about 3-7 barrels of low-sulfur (0.05 to 0.2 percent) synthetic oil from one ton of bituminous coal. The fuel oil has a heating value of 18,000 Btu/lb.

Solvent refining was initiated with the objective of producing a low-cost antipollution alternative to residual oil and natural gas. This process can produce either an ashless, low-sulfur solid product or a liquid fuel both with a heating value of about 15,900 Btu/lb. It is in the most advanced state of development of all of the coal liquefaction processes and was selected as the model process. The process involves adding hydrogen to the coal-solvent slurry and depolymerizing the coal in the reactor vessel. The sulfur is removed as hydrogen sulfide in the pressure let-down vessel, and the liquefied slurry is filtered, distilled, and solidified to produce the ashless solid product. A 2 tons of coal/hour pilot plant was started in mid-1974 at Fort Lewis, Washington. To date continuous integrated operation has not been achieved. In January, 1974, a 6 tons of coal/day pilot plant was started in Wilsonville, Alabama. It has operated intermittently at less than rated capacity.

The Consol process was designed to transform high-sulfur Eastern bituminous coal into a low-sulfur synthetic crude oil, or a fuel oil suitable for use in utility plants. The process involves slurrying the coal with the recycled solvent and heating to 750°F at about 400 psia; no hydrogen is added to the reactor, it is strictly a solubilization step.

The resulting slurry is passed through hydroclones to concentrate the solids. The overflow extract is hydrogenated over a Co-Mo catalyst and the solids are gasified to generate the required hydrogen. Consolidation Coal Company states that the Consol process is capable of producing 1.5 barrels of 0.2 percent sulfur fuel oil and 0.5 barrel of naphtha from one ton of coal feed.(38)

Environmental Problems

Environmental problems associated with coal liquefaction may involve significant health problems. It is well known that sufficient exposure to a variety of chemicals can cause cancer in man. Since 1900, it has been recognized that workers handling coal tars, certain aromatic amines, and some heavy metal compounds have increased incidence of carcinoma of the skin, bladder, and lungs, respectively. Likewise, other coal-derived products such as benzo(α)pyrene, dibenz(α)anthracite, 7, 12-dimethyl-benz(α)-anthracite and 3-methylchol-anthrene are known to be strong carcinogens. Therefore, prompt attention to conversion of waste to environmentally acceptable materials, hopefully at an economic advantage, is very important.

Applicability to NUC Sources

The applicability of liquefaction products to existing coal-, oil-, and gas-fired small industrial and commercial boilers will technically depend on boiler configuration and operation. Liquefaction products can be classified in terms of solid fuel such as solid SRC and liquid fuels such as liquid SRC, H-coal product, and other process products. According to the studies conducted by the Bureau of Mines,(39) Combustion Engineering,(40) and Babcock and Wilcox(41) as reported by Schreiber, et al.,(17) the solid SRC appeared similar to a high volatile bituminous coal except for the reduced sulfur and ash content. The grindability index, however, is high (about 16 percent) as compared with that of nonprocessed coal (about 60 for the Kentucky No. 11 coal). The liquid SRC was similar to No. 6 fuel oil in handling and combustion characteristics although the preheating requirement was greater. The liquid fuels obtained from other processes were quoted to be similar to a crude oil, fuel oil, or naphtha.(35-39)

According to the results of the industrial and commercial boiler data analysis conducted in this study, boiler subgroups of environmental concern are boilers burning high sulfur coal and fuel oil. Therefore, the application pattern of liquefied fuels to existing small boilers may be considered only for the coal- and oil-fired boilers. In general, conversions from oil-fired combustion equipment to coal fired are associated with an exorbitant cost. Only conversions from coal firing to oil firing are judged practical. Under these considerations, it was concluded that solid SRC can be applicable only to coal-fired boilers while liquid SRC, along with other liquid fuels of liquefaction process can be applicable to both coal- and oil-fired boilers. The use of liquid SRC, however, would require heating of all fuel handling equipment in contact with the fuel to above 350°F, resulting in a high boiler modification cost.

Model Plant Calculations

Of the various coal liquefaction processes under development, the SRC is the most advanced. In addition, it produces a solid form of fuel which can be readily used in the existing coal-fired boilers. For these reasons the process was chosen in the model plant calculation of liquefaction process. The liquid fuel which can be obtained from the process is difficult to handle in the existing coal- and oil-fired boilers, however, and is not considered to represent a typical liquid fuel from liquefaction. The H-coal process, therefore, was also treated in this analysis to examine the economics of liquid fuel application.

SRC Process

The SRC process is capable of reducing the sulfur and ash contents in the coal to 0.6 percent and 0.05 percent by weight, respectively. The heating value of the product is estimated at 15,900 Btu/lb regardless of the original heating value of the coal. H_2S gas is generated and utilized in a Claus reactor to produce elemental sulfur. When the SRC fuel is consumed in industrial and commercial boilers, the SO_2 emission is estimated at 0.75 $lb/10^6$ Btu.

Table 16 shows estimated costs for SRC manufacturing in mid-1973. The estimations were largely based on information provided by Battelle(42) and M. W. Kellogg Company.(43) The Utility Financing Method, as presented in Appendix A of this report, was employed. The estimated product cost is $\$1.04/10^6$ Btu or \$33.07/ton of product.

H-Coal Process

The estimated costs for the H-coal liquefaction process are shown in Table 17 for producing synthetic crude oil (Case 1) and for producing fuel oil and naphtha (Case 2). The estimations were based largely on the information provided by Hydrocarbon Research, Inc. The manufacturing cost was estimated at $\$1.38/10^6$ Btu and $\$1.34/10^6$ Btu for the production of synthetic crude oil and naphtha, respectively.

TABLE 16. COSTS FOR SOLVENT REFINED COAL PROCESS

Mid-1973, 8000 hours/yr Operation

Item	Quantity				
Product Rate (tons/day)	7,236	7,236	7,236	7,236	7,834
Type Coal	Eastern Medium S	Eastern High S	Central Medium S	Central High S	
Coal Input Rate (tons/day)	11,993	12,664	13,765	14,391	13,600
Sulfur Production Rate (long tons/day)	122	300	145	347	300
Capital Requirement $(10^6\$)$					
Coal preparation	8.18	8.50	9.01	9.30	7.29
Preheaters/dissolvers	22.17	23.03	24.42	25.20	29.15
Ash separation	10.81	11.42	12.41	12.98	10.93
Solvent/aromatics recovery	16.76	21.73	18.60	23.89	21.86
Sulfur recovery	4.44	7.29	4.89	7.90	7.29
Product handling	5.36	5.36	5.36	5.36	7.29
Hydrogen plant	8.54	8.87	9.40	9.70	7.29
Other	29.19	31.85	31.49	34.22	29.16
Total Bare Cost	105.45[a]	118.05[a]	115.58[a]	128.55[a]	120.26[b]
Engineering and design	5.27	5.90	5.78	6.43	6.01
Contractor fees	10.55	11.81	11.56	12.86	12.03
Subtotal Plant Investment	121.27	135.76	132.92	147.84	138.30
Project contingency	18.19	20.36	19.94	22.18	20.74
Total Plant Investment	139.46	156.12	152.86	170.02	159.04
Interest during construction	23.53	26.35	25.80	28.69	26.84
Startup	10.13	10.70	11.38	11.92	11.35
Working capital	10.13	10.70	11.38	11.92	11.35
Total Capital Requirement	183.25	203.87	201.42	222.55	208.58

TABLE 16. (Continued)

Item	Quantity				
Annual Costs (10^6\$/yr)					
Direct operating labor(c)	2.33	2.33	2.33	2.33	2.33
Maintenance (3.5 percent/yr)	4.88	5.46	5.35	5.95	5.57
Supervision	0.64	0.68	0.67	0.71	0.68
Administration and general overhead	2.95	3.11	3.08	3.25	3.15
Local taxes and insurance	3.77	4.22	4.13	4.59	4.29
Catalysts, chemicals, etc.(d)	0.50	0.50	0.50	0.50	0.50
Water(d)	0.20	0.21	0.23	0.24	0.23
Sulfur recovery supplies(e)	0.33	0.81	0.39	0.94	0.81
Gross Cost Excluding Coal	15.60	17.32	16.68	18.51	17.56
Aromatics and power credit(f)	-4.53	-5.03	-5.20	-5.74	-5.13
Sulfur credit ($10/long ton)	-0.41	-1.00	-0.48	-1.16	-1.00
Net Cost Excluding Coal	10.66	11.29	11.00	11.61	11.43
Coal cost ($10/ton)	39.98	42.21	45.88	47.97	45.33
Net Annual Operating Cost	50.64	53.50	56.88	59.58	56.76
Depreciation	8.66	9.66	9.50	10.53	9.86
Return on rate base	10.15	11.26	11.17	12.31	11.55
Federal income tax	3.35	3.71	3.68	4.06	3.81
Total Annual Cost	72.80	78.13	81.23	86.48	81.98
Product Cost ($/ton product)	30.18	32.39	33.68	35.85	31.39
Product Cost (\$/10^6 Btu output)(g)	0.95	1.02	1.06	1.13	0.99
Heat Output Rate (10^{12} Btu/yr)(g)	76.7	76.7	76.7	76.7	83.0

Footnotes to Table 16

(a) Bare cost of all sections except sulfur recovery based on Battelle Energy Program report "Liquefaction and Chemical Refining of Coal", July, 1974. Sulfur recovery section costs based on Shore, et al., EPA 650/2-74-098, September, 1974.(43)

(b) Bare cost of all sections based on Shore, et al.

(c) Operating labor = 175 men, consensus of two sources cited in (a).

(d) Based on Battelle Energy Program report.

(e) Based on Battelle analyses of requirements for amine scrubbing units and Claus plants. Total requirements per long ton of sulfur recovered by Claus plant are 92.5 kWh electricity, 13,300 lb steam, 54,000 gal cooling water, 1.81 lb monoethanolamine, 800 gal boiler feed water, and 0.4 lb activated alumina.

(f) Based on Battelle Energy Program report. Phenol at 1¢/lb, cresylic acids at 0.5¢/lb, power at 0.6¢/kWh, and 0.5 kWh per pound of ash burned.

(g) Based on 15,900 Btu/lb product.

TABLE 17. COSTS FOR H-COAL LIQUEFACTION PROCESS
Mid-1973, 8000 hours/yr Operation

Item	Quantity Case 1	Quantity Case 2
Products (bbl/day)		
Synthetic crude oil	67,466	
Fuel oil		48,122
Naphtha		15,000
By-products		
High Btu fuel gas (10^9 Btu/day)	89.90	34.09
Ammonia (tons/day)	205	152
Sulfur (long tons/day)	977	801
Capital Requirement (10^6\$)		
On-site investment	236.14	190.83
Off-site investment	41.45	33.41
Initial catalyst charge	2.59	1.51
Total Bare Cost	280.18	225.75
Engineering and design	14.01	11.29
Contractor fees	28.02	22.58
Subtotal Plant Investment	322.21	259.62
Project Contingency	48.32	38.93
Total Plant Investment[(a)]	370.53	298.55
Interest during construction	62.53	50.38
Startup	23.77	24.87
Working capital	23.77	24.87
Total Capital Requirement	480.60	398.67
Annual Costs (10^6\$/yr)[(b)]		
Payroll with benefits	4.46	4.46
Maintenance materials	5.36	4.31
Maintenance labor	6.38	4.82
Contracted services	0.60	0.60
Overhead and other expenses	2.00	2.00
Local taxes and insurance	10.00	8.06
Electricity (1¢/kWh)	16.78	13.11
Water (2¢/1000 gal)	0.11	0.08
Catalyst and chemicals	9.49	8.91
Gross Cost Excluding Coal	55.18	46.35

TABLE 17. (Continued)

Item	Quantity Case 1	Quantity Case 2
Annual Costs (10^6\$/yr)(b)		
Fuel gas credit (90¢/10^6 Btu)	-26.97	-10.23
Ammonia credit ($33/ton)	-2.26	-1.67
Sulfur credit ($10/long ton)	-3.26	-2.36
Net Cost Excluding Coal	22.69	31.78
Coal cost ($10/ton)	96.14	92.59
Net Annual Operating Cost	118.83	124.37
Depreciation	22.84	18.69
Return on rate base	26.48	22.24
Federal income tax	8.73	7.33
Total Annual Cost	176.88	172.63
Product Cost ($/bbl product)	7.87	8.20
Product Cost ($/$10^6$ Btu output)(c)	1.38	1.34
Heat Output Rate (10^{12} Btu/yr)(c)	127.9	128.8

(a) Total plant investment based on C. A. Johnson, et al., "Present Status of the H-Coal Process," Hydrocarbon Research, Inc., 1973.(35)

(b) Operating requirements and/or costs based on paper by C. A. Johnson, et al.

(c) Based on following densities and heating values:

Product	Density (°API)	Heating Value (Btu/lb)
Synthetic crude oil	25.2	18,000
Fuel oil	-3.1	16,700
Naphtha	50.0	18,700

Heating values from J. B. Maxwell, Data Book on Hydrocarbons, p 180, 1950.

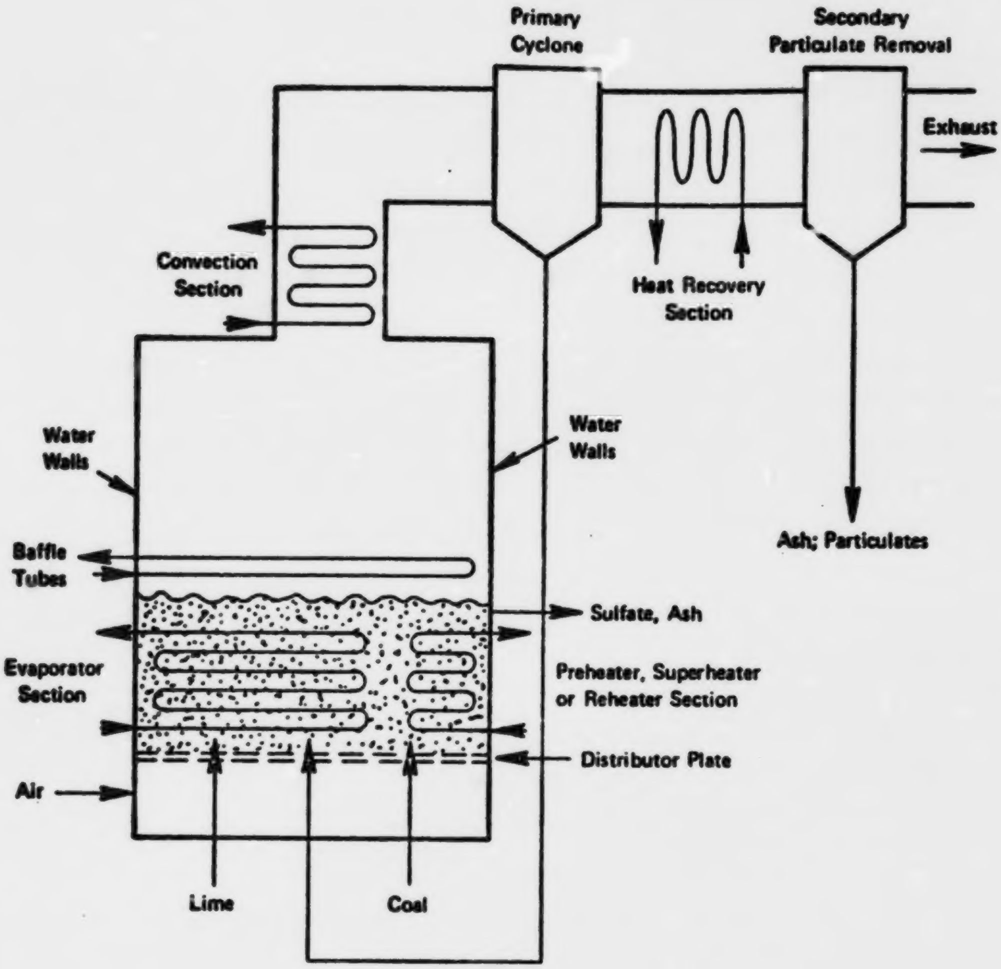
FLUIDIZED-BED COMBUSTION

One of the potentially viable techniques for SO_x control is fluidized-bed combustion (FBC) of high-sulfur fossil fuel. Winkler invented the fluidized-bed combustion concept in 1921 for use in coal gasification long before fluidized-bed technology came into general use in the 1940's for catalytic cracking in petroleum refining. Not until early in the 1960's, however, did fluidized-bed combustion as a boiler firing technique receive attention, first in Europe as a method of utilizing anthracite fines(44) and lignite.(45) In the early 60's, experimental programs on fluidized-bed combustion were undertaken by the National Coal Board (NCB)(46) and the British Coal Utilization Research Association (BCURA)(47,48) for the main purpose of reducing capital costs of power stations. In the United States, research programs were begun in the mid-60's by Pope, Evans, and Robbins (PER)(49) to develop packaged industrial boilers.

FBC Technology and Environmental Emissions

Figure 25 shows a simplified fluidized-bed combustion boiler concept. The combustion air passes through a bed of lime (or limestone), coal, and ash particles at such velocity (2-15 ft/sec) as to suspend all particles in the bed and to set all particles in a homogeneous fluid motion. In this state, the particles are separated from each other by an envelope of the fluidizing gas and present an extended surface for combustion. In addition, the randomly moving particles remain in the fluidized bed long enough for efficient combustion.

Fluidized-bed combustion has a high volumetric heat release rate of 500,000 $Btu/hr\text{-}ft^3$, as compared to 20,000 $Btu/hr\text{-}ft^3$ in a pulverized-coal-fired boiler. Also, the rapid movement of the solid particles passing over tubes immersed in the bed results in a high rate of heat transfer. Thus, smaller boilers with less tube surface should be possible for fluidized-bed combustion systems, allowing a 250,000 lb steam/hr industrial coal-fired boiler to be shop fabricated.


Primary Cyclone
Secondary Particulate Removal
Exhaust
Convection Section
Heat Recovery Section
Water Walls
Water Walls
Ash; Particulates
Baffle Tubes
Sulfate, Ash
Evaporator Section
Preheater, Superheater or Reheater Section
Distributor Plate
Air
Lime
Coal

Pressure: 1 – 25 atm
Coal Size: pf – 1/4 in.
Air Flow: 2 – 15 ft/sec
Temperature: 1400 – 1900°F

Surface: Water Walls, Horizontal, and Vertical Tubes in Bed
Sulfur removal: $CaO + SO_2 + \frac{1}{2}O_2 \rightarrow CaSO_4$

FIGURE 25. FLUIDIZED-BED COMBUSTION BOILER(44)

The pollutants from the process include residual emission of SO_2 and by-product solid wastes. As a sorbent material, dolomite is more efficient than limestone.(50) For a mole ratio of Ca/S of 3, dolomite was able to remove 95 to 99 percent while limestone was able to remove 75 to 85 percent of the total sulfur. This might be attributed to the fact that dolomite becomes more porous than limestone when carbonated. The by-product solid wastes in general contain ash, $CaSO_4$, CaO, unburned carbon, and MgO. The discharge rate depends on the Ca/S ratio and sorbent material employed. For a Ca/S ratio of 3, about 948 lb/ton of coal and 726 lb/ton of coal would be generated for using dolomite and limestone, respectively.

The fluidized-bed temperatures ($1400^{\circ}F$ to $1900^{\circ}F$) are selected to achieve the maximum SO_2 capture by the lime or limestone (over 90 percent removal). At these low temperatures, NO_x emissions are reduced (250-600 ppm) and clinkering problems are minimized. Experimental evidence indicates that the reaction of NO_x with CO to form N_2 is promoted by CaO.(51) Therefore, when combustion was carried out in two stages, one under reducing conditions (oxygen deficient) and one under oxidizing conditions (oxygen sufficient), NO_x emissions were reduced to 70 ppm. Pressurized operation also favors NO_x reduction (50-200 ppm NO_x at 5 atm).(52)

Applicability to NUC Sources

In FBC, the high volumetric heat-release rate and heat-transfer rate in the fluid-bed combustion system permits the design of compact boilers. Thus, an industrial boiler of up to 300,000 lb/hr steam capacity could be shop-fabricated and transported by rail. Design of FBC units for operation under pressure will have the effect of further reducing the size of the fluidized-bed boiler. It also enables a portion of the power to be generated by a gas-turbine yielding a higher overall thermal efficiency for the total plant. Thus, pressurized operation has advantages for larger boiler systems when it is used for the production of electrical power. This value is considerably decreased when the objective is process steam. The pressurized concept requires much more expensive components such as pressure units and particle control system, as compared to the atmospheric pressure design and this becomes a disadvantage for small plants. Therefore, a small pressurized fluidized-bed industrial boiler of capacity 250,000 lb/hr steam will not be economical compared to an atmospheric pressure unit.

Conceptually, FBC is a design for a new boiler system; most of the energy resulting from combustion is extracted by steam coils in the fluidized bed and only a small portion of the energy is carried over as sensible heat by the combustion flue gas. Steam coils are placed in the bed to control the bed temperature at $1400^{o}F$ to $1800^{o}F$ which is the optimum temperature for the reaction of CaO and SO_2.

Two conceptual approaches may be considered for the application to the existing industrial and commercial boilers. One of them is to install a fluidized-bed combustion unit prior to the existing boiler unit. Steam is generated both in the fluidized bed and the existing boiler unit. The existing boiler unit, however, is used as a heat exchanger to recover the sensible heat from the combustion gas. The net effect is that the existing facilities will be derated except for coal handling and storage facilities.

The other approach is to operate a FBC unit under reducing conditions wherein only a fraction of the stoichiometric amount of air necessary for complete combustion is employed in the fluidized bed. The temperature of the bed is maintained at about $1600^{o}F$ with a minimal amount of heat withdrawal through steam coils. The unburned carbon will be recycled to the carbon burnup cell where oxidizing conditions will be maintained to achieve complete combustion. A low-Btu gas (approximately 150 Btu/scf) resulting from the fluidized-bed combustion and a hot flue gas (at about $1600^{o}F$) resulting from the carbon burnup cell combustion are used in the existing boiler for steam generation. This type of combustion is similar to a low-Btu gasification process. This concept was not considered in this section of the study.

In retrofitting a FBC unit to an existing boiler system, modifications of the boiler configuration would be necessary. The pressure drop across the existing boiler system would be increased due to the increase in the flue gas flow rate. The pressure drop, however, may not exceed a certain level, i.e., 2 psi, due to the limited strength of the boiler wall and supporting system. Therefore, some of the baffles and steam coils should be removed to reduce the pressure drop. Moreover, the air heater or economizer may also have to be removed and, consequently, the boiler capacity and efficiency would be decreased.

Model Plant Calculation

In this study, two alternatives were analyzed to examine economic and technical feasibility. The first was to replace an existing boiler system with a new FBC boiler system and the second, to retrofit a FBC system to an existing boiler system. In the latter case, the FBC unit was assumed to be operated using excess combustion air, i.e., 150 percent. About 55 percent of the total steam was assumed to be generated from the existing boiler system and the remainder from the FBC unit.

The basis of model plant calculation is as follows:

- Steam generation capacity: 250×10^3 lb/hr
- Sulfur content of coal: 3 percent by weight
- Ash content of coal: 14 percent by weight
- Heating value of coal: 12,000 Btu/lb
- Fluidized-bed temperature: $1600^\circ F$
- Operating pressure: atmospheric
- Ca/S ratio: 3 by mole
- Sulfur removal scheme: once through.

The thermal efficiency of the adiabatic FBC unit may be assumed to be 90 percent excluding sensible heat loss from effluent gas stream.(53) The thermal efficiency of the existing boiler system when retrofitted with a FBC unit would be reduced to about 50 to 60 percent due to the boiler modification and the increased sensible heat loss from the effluent gas stream. Table 18 shows the operating conditions of the FBC systems under consideration. Table 19 shows estimated costs of two different FBC applications under consideration. Although the capital cost of the retrofit system is lower than that of the new system, the operating cost is higher due to the low thermal efficiency of the retrofit system.

TABLE 18. OPERATING CONDITIONS OF FBC SYSTEMS

Boiler Capacity = 250,000 lb steam/hr
Boiler Load Factor = 45 percent

Item	New FBC Boiler	Retrofit System
Excess Air, percent	10	150
Boiler Modification	None	The pressure drop may increase by about 3 times. To reduce gas flow resistance, baffles, if any, would be removed. If necessary, the air heater, economizer, or some of the steam coils would be removed. A bigger capacity fan should be employed.
Fraction of Steam Produced by Existing Boiler	0.0	0.55
Existing Boiler Efficiency, percent	N.A.	55
Overall Efficiency, percent	83	71
Overall Steam Generation Capacity, 10^3 lb/hr	250	250
Coal Requirement, 10^3 tons/yr	49.5	57.8
Limestone (or Dolomite) Requirement, 10^3 tons/yr	13.5 (24.6)*	15.8 (28.7)*
Solid Wastes, 10^3 tons/yr	18.0 (23.5)*	21.0 (27.4)*
Power Requirement, kW	300	500

* When dolomite is used instead of limestone.

TABLE 19. ESTIMATED COSTS OF FBC ALTERNATIVES FOR INDUSTRIAL BOILER APPLICATION

Item	New FBC Boiler	Retrofit FBC Boiler
Capital Requirement, $\$10^6$		
FBC System(a)	1.64	1.02
Boiler Modification(b)	0	0.10
Total Bare Cost	1.64	1.12
Engineering and Design	0.08	0.06
Contractor's Overhead and Profit	0.16	0.11
Subtotal Plant Investment	1.88	1.29
Project Contigency	0.28	0.19
Total Plant Investment	2.16	1.48
Interest During Construction(c)	0.10	0.07
Start-up Cost	0.08 (0.10)*	0.11 (0.14)
Working Capital	0.08 (0.10)	0.11 (0.14)
Capital Requirement	2.42 (2.46)	1.77 (1.83)
Retrofit	0	0.18 (0.18)
Total Capital Requirement	2.42 (2.46)	1.95 (2.01)
Operating Cost, $\$10^6$		
Labor(d)	0.06	0.05
Administrative and General Overhead	0.04	0.03
Materials and Utilities(e)	0.15 (0.22)	0.18 (0.27)
Solid Waste Disposal(f)	0.09 (0.12)	0.11 (0.15)
Additional Fuel Cost(g)	0.01	0.10
Local Taxes and Insurance	0.06	0.04
Gross Operating Cost	0.41 (0.51)	0.51 (0.64)
Credit	0	0
Net Operating Cost	0.41 (0.51)	0.51 (0.64)
Annualized Control Cost, $\$10^6$		
Return on Rate Base	0.13 (0.13)	0.11 (0.11)
Federal Income Tax	0.04 (0.04)	0.04 (0.04)
Depreciation	0.12 (0.12)	0.18 (0.19)
Net Operating Cost	0.41 (0.51)	0.51 (0.64)
Average Annual Cost	0.70 (0.80)	0.84 (0.98)
Annualized Control Cost,		
\$/lb S removed	0.29 (0.28)	0.30 (0.30)
$\$/10^6$ Btu Output	0.71 (0.81)	0.85 (0.99)

* Values in parentheses represent when dolomite is used instead of limestone.

Footnotes for Table 19

(a) A new FBC boiler system includes main FBC combustion unit, carbon burnup cell, superheater, economizer, air plenum, casing enclosure, ducts, structural supports, platforms, boiler trim, forced draft fan, auxiliary equipment, etc. It excludes coal preparation facilities and off-site facilities. The retrofitted FBC systems does not include carbon burnup cell, economizer, superheat, and air heater. Installation of steam coils in the retrofit FBC unit depends on the fraction of total steam to be generated from the unit. The barc cost of FBC system was estimated based on the information from References (53) and (54).

(b) Boiler modification cost includes costs for removing baffles and some of the steam coils, and cost for fan replacement. In the case of installing a new FBC boiler system, the dismantling cost of the existing boiler was assumed to be equal to the salvage value of the existing boiler system.

(c) Interest during construction was obtained by Interest During Construction = Total Plant Investment x Interest Rate (0.09) x Effective Construction Period (0.5 year).

(d) This includes the direct operating labor, maintenance, and supervision.

(e) This includes maintenance and operating supplies, limestone, or dolomite (at $7/ton) and power. The cost for fuel was not included.

(f) The solid waste disposal cost was assumed at $5/ton.

(g) The basis of boiler thermal efficiency was assumed to be 85 percent. The cost of coal was assumed at $10/ton.

FLUE GAS DESULFURIZATION (FGD) PROCESSES

Five processes are considered for post-combustion control of SO_2 from NUC sources: limestone, lime, double alkali, MgO (with both integrated and centralized regeneration), and Wellman-Lord. Below are brief process descriptions; more complete descriptions are included in Appendix B.

Process Descriptions

Limestone Slurry

The process considered here has been developed by Peabody Engineering.(55) Flue gas is first cleaned of particulate matter in an electrostatic precipitator (ESP) or equivalent. SO_2 is then reacted with $CaCO_3$ in a spray tower absorber where 70 to 90 percent of the sulfite is oxidized to sulfate. After vacuum filtration, the resulting 70 percent solids cake of $CaSO_3/CaSO_4$ is transported by truck to a landfill area. SO_2 removal efficiency is from 70 to 90 percent. Other than the $CaSO_3/CaSO_4$ (three pounds on a dry basis per pound of SO_2 removed) there is no waste produced. A full-scale (175MW) unit at Detroit Edison's St. Clair No. 6 is presently undergoing start-up.

Lime Scrubbing

A. B. Bahco Ventilation, Enkoping, Sweden has developed an industrial sized lime scrubbing process that is being marketed in the States by Research-Cottrell, Inc., Bound Brook, New Jersey. The Bahco process uses lime slurry in a two-stage venturi scrubber to remove particles and SO_2 from flue gas.(56, 57) Both $CaSO_3$ and $CaSO_4$ are produced and are removed from the process in the form of a sludge stream which is thickened and filtered. SO_2 removal ranges from 70 to 90 percent depending on the SO_2 concentration in the flue gas. Other than $CaSO_3/CaSO_4$ sludge (2.5 lb per pound of SO_2 removed on a dry basis) there are no waste streams. Currently, 19 commercial units have been installed in Japan and Sweden. Start-up of a 20 MW coal fired unit in the U. S. is expected in early 1976.

Double Alkali

FMC is one of a number of developers of processes that scrub with a Na_2SO_3 buffer solution and then react the clear solution with lime or limestone to precipitate $CaSO_3$.(58, 59) The purpose of separating scrubbing from precipitation has been to eliminate scaling difficulties. Since the sludge removed from the system contains four to five percent Na_2SO_3 and Na_2SO_4, soda ash must be added to the system to replace these sodium losses. Removal efficiencies have been 99 percent for flyash and 90 percent for SO_2. Start-up of a 45 MW unit is expected shortly.

MgO Process

Chemical Construction Corporation, New York, New York, has developed a regenerable FGD process that has eliminated sludge disposal problems.(60-62) Flue gas passes through an ESP and contacts a finely divided slurry of MgO in a venturi scrubber. SO_2 reacts with the MgO to form hydrated $MgSO_3$ and a small amount of $MgSO_4$. The $MgSO_3$ and $MgSO_4$ are centrifuged and dried. Subsequently, the dried product is taken to a regeneration facility where it is calcined forming MgO and driving off SO_2 which can be used to produce high grade sulfuric acid. Ninety percent SO_2 removal efficiency has been demonstrated. In addition to make-up MgO, fuel oil is required to fire the calciner and the drier. The process has been demonstrated on a 155 MW oil fired boiler (Boston Edison's Mystic Station) and on a 190 MW coal fired boiler (Potomic Electric Power's Dickerson No. 4).

Wellman-Lord

In this process, flue gas which has been cleaned of particulate matter is contacted with a slurry of Na_2SO_3, $NaHSO_3$, and Na_2SO_4. SO_2 reacts with the sulfite to form bisulfite.(63) In addition, some oxidation of sulfite takes place. A ten percent slip stream is sent to an evaporator where the bisulfate is decomposed, regenerating the sulfite and evolving SO_2 which can be used in the production of H_2SO_4 or elemental sulfur. A certain amount of sulfate and thiosulfate is produced in the bisulfate decomposition. These species are treated in a purge treatment system. NaOH or Na_2CO_3 are added to make up for sodium ion lost in the purge. SO_2 removal has been 90 percent. In addition to sodium make-up, steam is required for evaporator operation. A large number of units are presently operating in Japan on acid plants and oil-fired boilers. A coal-fired demonstration at Northern Indiana Public Service Corporation's Gary, Indiana power plant is scheduled for start-up in January 1976.

Applicability to NUC Sources

Although most of the flue gas desulfurization (FGD) systems applicable to utility boilers can be used on NUC sources, the application to NUC sources differs somewhat from its application to utility boilers. Many NUC sources release stack gas at 400 to 500°F, so there is greater potential for gas reheat by heat recovery from the incoming gas. In general the NUC sources require larger excess air than the utility boilers. Because of larger requirements for excess air, SO_2 concentration will be lower and oxygen concentration will be higher. This can possibly cause difficulties in processes where oxidation is undesirable such as the double alkali and Wellman-Lord processes. Higher excess air also means that a larger quantity of flue gas must be handled for a given quantity of fuel. On the other hand, the reduced amount of particulate matter in the flue gas could mean less operating problems than have been experienced at utility FGD systems.

The size of control systems for NUC sources permits more shop fabrication of system components than for utility applications. Unlike utilities, industrial diversity may also permit unusual system configurations, especially where there are captive uses for sulfur compounds or in-house capabilities for regeneration. For example, the pulp and paper industry can regenerate sodium-sulfite scrubbing solutions in its manufacturing operation. The chemical industry frequently has captive uses for sulfuric acid or sulfate salts. The petroleum industry has sources of hydrogen sulfide that could be used to produce sulfur from SO_2 emissions. Such special cases will be more abundant in industrial applications than in utility applications.

The turndown ratio for an NUC source is such that the boiler is shut down and started up many times during a year. Therefore, an FGD system must have load following capability. This type of operation can be achieved by a high degree of automation and by providing a large surge volume after the scrubber so that the regeneration or waste disposal system can continue to operate when the boiler and scrubber are shut down. Of course, remote regeneration or waste disposal facilities tied to several boilers would automatically provide the surge volume.

The relatively small size of NUC sources may offer some unique situations for the disposal of purge streams or waste products from FGD systems. The volume of the purge stream or the tonnage of the waste products can be two orders of magnitude less than for a typical utility boiler. Water authorities may allow the purge stream to be discharged to city sewer system when combined with other waste streams in the plant. The waste products such as calcium sulfite/sulfate sludge may be trucked to a nearby sanitary landfill for disposal.

Industrial plants in general are built with a higher ratio of equity/debt as compared with utilities; thus, taxes are higher. Also, nonregulated industry requires higher return on investment because of the risk involved. The combined effect results in a higher annualized capital charge. Moreover, the relative impact of capital costs is increased because of the small scale of operation. In addition, the annualized cost per unit of heat output can be high because of the low load factor for NUC sources.

Model Plant Calculation

To determine the control cost of each of the FGD processes described above, a conceptual cost study was carried out for a 250,000 lb steam/hour boiler. The coal was assumed to contain 3 percent of sulfur, 14 percent of ash, and 12,000 Btu/lb of heating value. The labor, materials, and utility requirements for each process are shown in Table 20 and the estimated control costs are shown in Table 21. The Utility Financing Method, as presented in Appendix A of this report, was employed for the estimations.

TABLE 20. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR FGD PROCESSES

Capacity: 250,000 lb steam/hr (65,600 scfm flue gas)
Coal: 3% sulfur; 14% ash; 12,000 Btu/lb heating value

Item	Peabody Limestone	Bahco Lime	FMC Double Alkali	MgO Integrated	MgO Central Regeneration	Wellman-Lord
Utility						
Power, kW	550	470	560	500	350	800
Steam, lb/hr	4,700	4,700	6,000	0	0	9,300
Water, gal/min	80	32	39	344	28	270
Fuel Oil, gal/hr	0	0	0	71	39	0
Materials						
Lime, ton/hr	0	0.74	0.91	0	0	0
Limestone, ton/hr	1.5	0	0	0	0	0
Soda Ash, ton/hr	0	0	0.14	0	0	0.037
Natural Gas, scf/hr	0	0	0	0	0	6,300
MgO, ton/hr	0	0	0	0.009	0.009	0
Coke, ton/hr	0	0	0	0.008	0	0
Maintenance, % TPI*/yr	2.0	1.5	1.0	3.5	2.0	2.0
Labor						
Operation, man/shift	1	0.5	0.5	2	1	2
Maintenance, % TPI/yr	1.0	1.5	1.0	3.5	2.0	2.0

* TPI indicates total plant investment.

TABLE 21. ESTIMATED CONTROL COST FOR FGD PROCESSES

Boiler Capacity: 250,000 lb steam/hr (65,600 scfm flue gas)
Load Factor: 45 percent
Coal: 3% sulfur; 14% ash; 12,000 Btu/lb heating value

	Cost (mid-1973), 10^3 dollars					
Item	Peabody Limestone	Bahco Lime	FMC Double Alkali	MgO-Integrated	MgO-Central Regeneration	Wellman-Lord
Capital Requirement						
Bare cost	1,452(a)	--	1,962(l)	2,481(q)	1,428(u)	1,995(y)
Engineering and design	--	--	--	--	--	--
Contractor's overhead and profit	138	--	--	--	--	199
Subtotal Plant Investment	1,590	--	1,962	2,481	1,428	2,194
Project Contingency	239	--	264	--	--	329
Total Plant Investment	1,829	--	2,226	2,481	1,428	2,523
Interest during construction	82	--	100	112	64	114
Startup cost	54	--	71	80	34	62
Working capital	54	--	71	80	34	62
Capital Requirement	2,019	2,520(g)	2,468	2,753	1,560	2,761
Capital requirement for boiler retrofit and modification(b)	606	756	494	826	468	828
Total Capital Requirement	2,625	3,276	2,962	3,579	2,028	3,589
Annual Operating Cost						
Labor	30(c)	30(h)	23(m)	102(r)	38(v)	69(z)
Administrative and general overhead	18	18	14	61	23	41
Materials and utilities	125(d)	129(i)	180(n)	168(s)	61(w)	132(aa)
Solid waste disposal or central regeneration	46(e)	63(j)	77(o)	0	11(x)	10(bb)
Additional fuel requirement	0	0	0	0	0	--
Local taxes and insurance	49	62	60	67	39	56
Gross Operating Cost	268	302	354	398	172	308
Credit	0	0	0	69(t)	0	13
Net Operating Cost	268	302	354	329	172	295
Annualized Control Cost						
Return on rate base	141	175	159	192	108	192
Federal income tax	46	58	53	63	36	63
Depreciation	257	322	290	350	200	352
Net operating cost	268	302	354	329	172	295
Average Annual Cost	712	857	856	934	516	902
Annualized Control Cost						
$/lb S removed	0.34(f)	0.35(k)	0.33(p)	0.36	0.20	0.35
\$/10^6 Btu output	0.79	0.87	0.87	0.95	0.52	0.92

Footnotes to Table 21

(a) The exponent scale-up factor was assumed to be 0.5.

(b) The retrofit factor was assumed at 1.3 for Peabody limestone, Bahco lime, MgO-integrated, MgO-central regeneration, and Wellman-Lord processes. The retrofit factor was assumed at 1.2 for FMC double alkali process.

(c) Direct operating labor = 1 man/shift.
Maintenance = 1/3 man/shift.

(d)

Limestone:	$131,000	
Power:	48,000	
Steam:	20,000	
Water:	22,000	
Maintenance:	37,000	(2 percent of total plant investment)
Supply:	20,000	(3 percent of labor)
	$278,000	at full load

(e) Solid waste generation = 4 lb/lb of SO_2 removed.

(f) Coal required at full load = 1.07×10^5 tons/year;
Sulfur removed at 80 percent efficiency = 2,577 tons/year.

(g) $(\$2.3 \text{ million}) \frac{(144.1)}{164.7} = \$2.01 \text{ million}$, where 144.1 and 164.7 are CE plant cost indexes for 1973 and 1974, respectively.

$$(\$2.01 \text{ million})\left(\frac{565{,}000}{450{,}000}\right)^{0.6} = \$2.52 \text{ million.}$$

This represents the capital requirement.

(h)

Direct operation	$ 22,000
Maintenance:	35,000
Supervision:	9,000
	$ 66,000 at full load

(i)

Operation supplies:	$ 20,000
Maintenance:	35,000
Power:	41,000
Water:	8,000
Reheat steam:	20,000
Lime:	165,000
	$286,000 at full load

(j) Sludge generated = 5.1 lb/lb of SO_2 removed.

(k) Sulfur removed = 2.738 tons/year.

Footnotes to Table 21 (continued)

(l) The bare cost of a 45 MW system(59) = $2.791 x 10^6. The bare cost includes engineering and design and contractor's overhead and profit. The exponent scale-up factor = 0.6.

(m)

Direct operation:	$ 22,000	
Maintenance:	22,000	(1 percent of total plant investment)
Supervision:	7,000	
	$ 52,000	at full load

(n)

Lime:	$200,000	
Soda ash:	61,000	
Power:	49,000	
Steam:	43,000	
Water:	10,000	
Operating supplies:	16,000	
Maintenance:	22,000	(1 percent of total plant investment)
	$401,000	at full load

(o) Solid wastes generated = 5.85 lb/lb of SO_2 removed.

(p) Sulfur removed = 2.899 tons/year at full load.

(q) The bare cost for a 200 MW boiler in 1972 = $11.476 x 10^6 (estimated from Reference 61). The bare cost for a 25 MW system in mid-1973 =

$$(11.476 \times 10^6) \left(\frac{100}{200}\right)^{0.8} \left(\frac{25}{100}\right)^{0.6} \left(\frac{144.1}{137.2}\right) - 0.532 \times 10^6 \text{ (for ESP)} =$$

$2.481 x 10^6. This cost includes costs for engineering and design, contractor's overhead and profit, and project contingency.

(r)

Direct operation:	$110,000	
Maintenance:	87,000	(3.5 percent of TPI)
Supervision:	30,000	
	$227,000	at full load

(s)

Power:	$ 44,000	
Fuel oil:	66,000	
Process water:	90,000	
MgO:	16,000	
Coke:	2,000	
Maintenance:	87,000	(3.5 percent of TPI)
Operating supplies:	68,000	
	$373,000	at full load

Footnotes to Table 21 (continued)

(t) SO$_2$ removed at full load = 5.798 tons/year;
S$_2$SO$_4$ (98 percent) = (1.33)(5.798) = 7.711 tons/year.

(u) Bare cost for an integrated MgO system = \$11.476 x 10^6 [from Footnote (q)]
Bare cost for the scrubbing system only =

$$(11.476 \times 10^6)\left(\frac{8.52}{13.1}\right)\left(\frac{100}{200}\right)^{0.8}\left(\frac{25}{100}\right)^{0.6}\left(\frac{144.1}{137.2}\right) - 532{,}000 \text{ (for ESP)}$$

= \$1.428 x 10^6 (total plant investment).

(v) Direct operation: \$ 44,000
Maintenance: 29,000 (2 percent of TPI)
Supervision: 11,000

\$ 84,000 at full load

(w) Maintenance: \$ 29,000
Operating supplies: 25,000
Power: 26,000
Fuel oil: 47,000
Water: 9,000

\$136,000 at full load

(x) The regeneration cost was estimated at \$8.53/ton sulfur removed (see Appendix C).

(y) Total bare cost of a Wellman-Lord system handling a flue gas of 294,000 scfm = \$5.856 x 10^6 (Battelle's estimate in 1974). Total bare cost of a Wellman-Lord system installed on a 250,000 lb/hr capacity boiler

$$= (5.856 \times 10^6)\left(\frac{65{,}600}{294{,}000}\right)^{0.6}\left(\frac{144.1}{172.0}\right) = \$1.995 \times 10^6 \text{ (in mid-1973).}$$

(z) Direct labor and supervision: \$113,000
Maintenance: 41,000

\$154,000 at full load

(aa) Power: \$ 20,000
Process water: 7,000
Cooling water: 32,000
Steam: 71,000
NaOH: 75,000
Natural gas: 50,000
Maintenance: 41,000

\$296,000 at full load

(bb) Waste disposal cost = \$22,800 at full load.

EVALUATION OF ALTERNATIVES

Approach

The evaluation of the potential role of the various alternatives under consideration in the control of the emissions from small industrial and commercial boilers requires consideration of a number of diverse factors which must be related and compared in a meaningful fashion. The approach involves the following steps:

(1) Development of evaluation criteria
(2) Evaluation of each alternative with respect to each criterion.

The conversion of the evaluation to a rating scale would be desired for the rating of the alternatives based on the aggregate points. However, the procedure involves subjective judgments which would influence the outcome significantly. The quantitative analysis of the evaluation, therefore, was not conducted in this study.

Evaluation Criteria

A set of six criteria is employed in the evaluation of the alternatives as follows:

(1) Pollutant emissions
(2) Retrofitability
(3) Operation maintenance
(4) Capital requirement
(5) Annualized cost
(6) Availability.

The alternatives under consideration have differing potential for minimizing air pollutant emissions and generating new pollutant emissions. The variability is expressed in terms of residual and secondary emissions which result from the application of an alternative.

The application of the alternatives to the existing industrial boiler systems should be made relatively easily. The variability is evaluated with respect to space requirement and boiler modification.

Alternatives employed are to be operated and maintained as trouble-free as possible. Operation-maintenance is evaluated with respect to material handling, technical expertise, the number of moving parts, plugging and scaling possibilities, corrosion and erosion possibilities, and operating temperature and pressure.

Capital requirement indicates the amount of capital required to incorporate an alternative process. The contribution of capital cost to annual operating cost is included in the annualized cost.

Annualized cost consists of return on rate base, Federal income tax, depreciation, and net annual operating cost.

In view of the urgency of related environmental problems, the availability of given alternatives is an important criterion in the evaluation. Factors such as raw material availability, developmental status, year of commercialization, and growth rate are components of the availability consideration.

Other factors such as by-product were not established as separate criteria since the criterion would not be a significant factor for small boilers. Besides, the factor is incorporated in the annualized cost as credit.

Alternative Evaluation

The next step in the procedure was to develop an evaluation of each alternative with respect to each of the six criteria. The evaluation was carried out based on the boiler operation viewpoint. That is, in the determination of pollutant emissions of the alternatives, the quantity was limited to the emissions resulting from the combustion of the fuel in boiler but did not include the emissions from the fuel conversion process unless the process was assumed to be retrofitted to the boiler system. Similarly, in the evaluation with respect to operation-maintenance, only the difficulties that might take place at the site of boiler operation were considered for evaluation. Therefore, the alternatives of fuel substitution were considered to be free of operation-maintenance problems.

A quantitative evaluation was employed wherever possible, otherwise qualitative categories for evaluation were developed. The evaluation of alternatives with respect to capital requirement and annualized cost were not included in this section but are discussed extensively in the next section since the relative costs depended on the size and operating characteristics of the boilers.

Pollutant Emission

The residual emission of sulfur dioxide was evaluated in terms of pounds of sulfur dioxide per million Btu steam output. The secondary emissions resulting from the sulfur dioxide control process were expressed in terms of the quantity of pollutants per million Btu of steam output. available.

Retrofitability

The retrofitability was evaluated on the basis of space requirement and need for boiler modification. The space requirement was categorized by four groups as follows:

Category 1 - No space requirement
Category 2 - Low space requirement
Category 3 - Moderate space requirement
Category 4 - High space requirement.

The need for boiler modification was evaluated with respect to four categories.

Category 1 - Need for no modification
Category 2 - Need for low modification
Category 3 - Need for moderate modification
Category 4 - Need for high modification.

Operation-Maintenance

The operational and maintenance difficulties were assessed in terms of technical expertise; characteristics of material handling; possibilities of plugging, scaling, erosion, and corrosion; number of moving parts, and operating pressures and temperatures.

The degree of technical expertise required is based on the process complexity, control sensitivity, and operating conditions. It was expressed in terms of technical knowledge equivalent to one of either technician or engineer.

The material handling was evaluated in terms of gases, liquids, solids, and slurry. The handling of solids or slurry is more difficult than that of liquids or gases.

Some sorption processes involve solid or slurry streams that are more susceptible to scaling and/or plugging than others. Scaling and plugging can precipitate equipment failure and result in operation disruptions. The potential was evaluated with respect to three categories as follows:

Category 1 - Minimal possibility
Category 2 - Moderate possibility
Category 3 - High possibility.

In SO_2 sorption processes, corrosion is caused primarily by the presence of dilute sulfuric acid and/or chlorine ions. Erosion is caused by the abrasive nature of liquids and solids. Both corrosion and erosion were evaluated with respect to three categories.

Category 1 - Minimal possibility.
Category 2 - Moderate possibility.
Category 3 - High possibility.

The number of moving parts was the summation of all of the major pieces of equipment containing moving parts. This included conveyors, rotary drum filters, pumps, blowers and mixers. This factor was categorized in terms of low, moderate, and high.

Operating pressures and temperatures influence the reliability of process operation to some extent. A high operating pressure or temperature is more conducive to failure than a low one.

Availability

The availability was evaluated on the basis of raw material availability, development status, and the year of commercialization and growth rate (i.e., the rate of implementation). The raw material availability was evaluated on the basis of two categories defined as follows:

Category 1 - Materials readily available and in surplus generally through the United States

Category 2 - Materials either in short supply or available only to specific areas.

The development status was classified into five categories--conceptual, bench, pilot, prototype, and commercial. The commercialization was evaluated in terms of the estimated year of commercial availability as applied to industrial boiler systems. A major factor influencing the rate of implementation is the complexity of the alternative. A highly complex process, requiring a longer lead time for fabrication of components and construction, and being more capital intensive will result in a lower implementation rate. With these considerations, the alternatives were evaluated with respect to three categories defined as follows:

Category 1 - Low degree of complexity

Category 2 - Intermediate degree of complexity

Category 3 - Highly complex process.

Evaluation Result

The results of the evaluation based on the criteria discussed above are shown in Tables 22 and 23 for coal- and oil-fired boilers,

TABLE 22. ALTERNATIVE EVALUATION MATRIX FOR COAL-FIRED BOILER

Alternative	Pollutant Emission		Retrofitability		Operation - Maintenance					
	SO_2 (lb/10^6 Btu)	Others (lb/10^6 Btu)	Space Requirement	Boiler Modification	Technical Expertise	Material Handling	Plugging Scaling	Erosion Corrosion	Moving Parts	Operating Condition
Physical Cleaning	NA*	None	None	None	← Normal Operation →					
Coal Gasification (Low-Btu)	0.52	None	High	Moderate	Engr	Liquid	Minimum	Moderate	High	High Temp
Coal Gasification (High-Btu)	None	None	None	Moderate	← Normal Operation →					
Coal Liquefaction (Solid SCR)	0.75	None	None	Low	← Normal Operation →					
Coal Liquefaction (Liquid SRC)	0.75	None	None	High	← Normal Operation (Heating of Fuel Required) →					
Coal Liquefaction (H-Coal)	0.22	None	None	Moderate	← Normal Operation →					
Fluidized Bed Combustion	1.00	solid waste (42.6)	Low	High	Tech	Solid	Moderate	Moderate	Low	High Temp
Limestone Slurry (Peabody)	1.00	sludge (16.0)	Medium	None	Tech	Slurry	High	Moderate	Moderate	Normal
Lime Scrubbing (Bahco)	0.75	sludge (21.7)	Medium	None	Tech	Slurry	High	Moderate	Moderate	Normal
Double Alkali (FMC)	0.50	sludge (26.3)	Low	None	Tech	Liquid	Minimum	Low	Moderate	Normal
MgO (Integrated)	0.50	None	High	None	Engr	Slurry	Moderate	High	High	High Temp
MgO (Central Regeneration)	0.50	None	Low	None	Tech	Slurry	Moderate	Moderate	Moderate	High Temp
Wellman-Lord	0.50	Purge Stream (2.5)	High	None	Engr	Liquid	Minimum	Moderate	High	High Temp

* NA = Not Applicable.

TABLE 22. ALTERNATIVE EVALUATION MATRIX FOR COAL-FIRED BOILER (Continued)

Alternative	Availability			
	Raw Material Availability	Developmental Status	Commercialization	Growth Rate
Physical Cleaning	Available	Commercial	Present	High
Coal Gasification (Low-Btu)	Available	Commercial	Present	Medium
Coal Gasification (High-Btu)	Available	Prototype	1980-1983	Low
Coal Liquefaction (Solid SCR)	Available	Prototype	1981	Low
Coal Liquefaction (Liquid SRC)	Available	Prototype	1981	Low
Coal Liquefaction (H-Coal)	Available	Pilot	1983	Low
Fluidized Bed Combustion	Available	Conceptual (Proven Tech)	1978-1980	High
Limestone Slurry (Peabody)	Available	Commercial	Present	High
Lime Scrubbing (Bahco)	Available	Commercial	Present	High
Double Alkali (FMC)	Available	Commercial	Present	High
MgO (Integrated)	Available	Commercial	Present	Medium
MgO (Central Regeneration)	Available	Commercial	Present	High
Wellman-Lord	Questionable	Commercial	Present	Medium

TABLE 23. ALTERNATIVE EVALUATION MATRIX FOR OIL-FIRED BOILER

Alternative	Pollution Emission		Retrofitability		Operation - Maintenance					
	SO_2 ($lb/10^6$ Btu)	Other ($lb/10^6$ Btu)	Space Requirement	Boiler Modification	Technical Expertise	Material Handling	Plugging Scaling	Erosion Corrosion	Moving Parts	Operating Condition
Coal Gasification (High-Btu)	None	None	None	Moderate	← Normal Operation →					
Coal Liquefaction (Liquid SRC)	0.75	None	None	Moderate	← Normal Operation (Heating of Fuel Required) →					
Coal Liquefaction (H-Coal)	0.22	None	None	None	← Normal Operation →					
Limestone Slurry (Peabody)	0.52	sludge (0.83)	Medium	None	Tech	Slurry	High	Moderate	Moderate	Normal
Lime Scrubbing (Bahco)	0.39	sludge (11.2)	Medium	None	Tech	Slurry	High	Moderate	Moderate	Normal
Double Alkali (FMC)	0.26	sludge (13.6)	Low	None	Tech	Liquid	Minimum	Low	Moderate	Normal
MgO (Integrated)	0.26	None	High	None	Engr	Slurry	Moderate	High	High	High Temp
Mgo (Central Regeneration)	0.26	None	Low	None	Tech	Slurry	Moderate	Moderate	Moderate	High Temp
Wellman-Lord	0.26	Purge Stream (0.7)	High	None	Engr	Liquid	Minimum	Moderate	High	High Temp

TABLE 23. ALTERNATIVE EVALUATION MATRIX FOR OIL-FIRED BOILER (Continued)

Alternative	Availability			
	Availability	Developmental Status	Commercilization	Growth Rate
Coal Gasification (High-Btu)	Available	Prototype	1980-1983	Low
Coal Liquefaction (Liquid SRC)	Available	Prototype	1981	Low
Coal Liquefaction (H-Coal)	Available	Pilot	1983	Low
Limestone Slurry (Peabody)	Available	Commercial	Present	High
Lime Scrubbing (Bahco)	Available	Commercial	Present	High
Double Alkali (FMC)	Available	Commercial	Present	High
MgO (Integrated)	Available	Commercial	Present	Medium
Mgo (Central Regeneration)	Available	Commercial	Present	High
Wellman-Lord	Questionable	Commercial	Present	Medium

respectively. Oil-fired boilers in general have fewer alternatives than coal-fired boilers due to the unique boiler configurations which do not lend themselves to solid fuels and low-Btu gases. The FGD processes employed for the evaluation were considered commercially available for the application to commercial and industrial boilers because of the small size. The type of fuel tested was not taken into consideration here.

COST OF ALTERNATIVES

This section is concerned with the control cost of the alternatives under consideration when applied to NUC sources such as small commercial and industrial boilers. According to the results of the boiler data analysis conducted in this study, the area source SO_2 emissions appear to be concentrated in high sulfur coal-fired boilers of between 10,000 and 500,000 lb steam/hr size class and in high sulfur oil-fired boilers of sizes between 1,000 and 500,000 lb steam/hr. For the purpose of conducting the control cost analysis, two different boiler size subgroups for coal-fired boilers and three different boiler size subgroups for oil-fired boilers were selected. The characteristics of the selected boiler subgroups are shown in Tables 24 and 25 for coal- and oil-fired boilers, respectively. For convenience of analysis, the nominal standard properties of the fuels employed in the cost estimation were assumed as shown in Table 26.

The total capital requirements of the alternatives consisted of costs for on-site installed facilities and costs for retrofit and boiler modification. It was a battery limit cost otherwise specified and included costs for equipment, materials, installation, engineering and design, and startup. Credits for any existing facilities were incorporated in the estimation. The base year for the cost estimates was mid-1973 and the Utility Financing Method(27) presented in Appendix A of this report was employed in the estimation of the related costs.

The annualized cost in general consisted of capital charges, maintenance, labor, utilities, raw materials, by-product credits, and additional costs or credits due to the use of the control alternatives.

The estimation of control cost for the processed fuel alternatives such as coal cleaning, gasification, and liquefaction processes was somewhat different in procedure from that for retrofit control systems such as FGD processes. Tables 27 through 32 show the estimated control cost of the processed fuel alternatives as applied to the selected boiler subgroups described above.

The control cost of the FGD processes was estimated based on the same format used in the FGD sections. The following assumptions were

TABLE 24. CHARACTERISTICS OF SELECTED COAL-FIRED COMMERCIAL AND INDUSTRIAL BOILERS

Item	Commercial (10^3 lb steam/hr)		Industrial (10^3 lb steam/hr)	
	20	250	20	250
Load factor	0.42	0.31	0.55	0.45
Excess combustion air, percent				
Pulverized coal	60*	50	60*	50
Stoker	100*	100	100*	100
Flue gas flow rate, scfm				
Pulverized coal	5,300*	62,500	5,300*	62,500
Stoker	6,700*	68,800	6,700*	68,800
Boiler efficiency,** percent	85	85	85	85

* Battelle's estimate based on pertinent information.

** Nominal value.

TABLE 25. CHARACTERISTICS OF SELECTED OIL-FIRED COMMERCIAL AND INDUSTRIAL BOILERS

Item	Commercial (10^3 lb steam/hr)			Industrial (10^3 lb steam/hr)		
	2	20	250	2	20	250
Load factor	0.40	0.28	0.19	0.45	0.36	0.41
Excess combustion air, percent	36	33	52	36	33	52
Flue gas flow rate, scfm	450	4,430	63,300	450	4,430	63,300
Boiler efficiency,* percent	85	85	85	85	85	85

* Nominal value.

TABLE 26. NOMINAL STANDARD PROPERTIES OF FUEL

Property	Coal	Oil
Sulfur content, weight percent	3	2.3
Ash content, weight percent	14	—
Heating value	12,000 Btu/lb	6×10^6 Btu/bbl

TABLE 27. CONTROL COST ANALYSIS FOR PHYSICAL COAL CLEANING

Item	Coal-Fired Boiler, 10^3 lb/hr Steam[c]				Oil-Fired Boiler, 10^3 lb/hr Steam					
	20(C)	20(I)	250(C)	250(I)	2(C)	2(I)	20(C)	20(I)	250(C)	250(I)
Boiler Load Factor (%)	42	55	31	45	40	45	28	36	19	41
Flue Gas Flow Rate (scfm)	6,000	6,000	65,600	65,600	500	500	4,400	4,400	63,300	63,300
Total Capital Requirement for Boiler Modification (10^3 $)										
Annual Operating Cost (10^3 $/yr)										
Fuel Cost[a]	47.0	61.6	434	630						
Less Base Case Fuel Cost	-36.1	-47.2	-333	-483						
Investment-Related Cost										
Effect on Boiler Operating Cost										
Total	10.9	14.4	101	147						
Sulfur Removed (10^6 lb/yr)[b]	0.078	0.102	0.719	1.043						
Control Cost ($/lb S Removed)	← 0.141 →									
Control Cost ($/$10^6$ Btu Output)	← 0.149 →									

(a) 54.3 ¢/10^6 Btu, based on analysis of 500 ton/hr plant.

(b) Based on cleaned coal heating value of 12,500 Btu/lb and sulfur content of 2 weight percent.

(c) C and I indicate commercial and industrial boilers, respectively.

TABLE 28. CONTROL COST ANALYSIS FOR KOPPERS-TOTZEK COAL GASIFICATION

Item	Coal-Fired Boiler, 10^3 lb/hr steam[f] 20(C)	20(I)	250(C)	250(I)
Boiler Load Factor (%)	42	55	31	45
Flue Gas Flow Rate (scfm)	6,000	6,000	65,600	65,600
Total Capital Requirement for Boiler Modification (10^3 \$)[a]	29.7	29.7	153	153
Annual Operating Cost (10^3\$/yr)				
Fuel Cost[b]	102.2	133.8	942	1,368
Less Base Case Fuel Cost	-36.1	-47.2	-333	-483
Investment-Related Cost[c]	4.3	4.3	22.3	22.3
Effect on Boiler Operating Cost[d]	-14.0	-18.3	-129	-187
Total	56.4	72.6	502.3	720.3
Sulfur Removed (10^6 lb/yr)[e]	0.197	0.258	1.82	2.64
Control Cost (\$/lb S Removed)	0.286	0.281	0.276	0.273
Control Cost (\$/10^6 Btu Output)	0.77	0.75	0.74	0.73

(a) Based on R. Schreiber, et al., EPA-650/2-74-123, November, 1974, Section 6.

(b) \$1.18/10^6 Btu.

(c) 14.55 percent of investment per year, based on utility financing method with no maintenance.

(d) Based on average 1973 difference in operating cost (excluding fuel) between gas-fired and coal-fired boilers (-0.19 mills/kWh). Values from Electrical World, November 1, 1973.

(e) Difference between sulfur emission for base case and sulfur emission at boiler and coal gasification plant.

(f) C and I indicate commercial and industrial boilers, respectively.

TABLE 29. CONTROL COST ANALYSIS FOR HYGAS COAL GASIFICATION

Item	Coal-Fired Boiler, 10^3 lb/hr Steam[f]				Oil-Fired Boiler, 10^3 lb/hr Steam					
	20(C)	20(I)	250(C)	250(I)	2(C)	2(I)	20(C)	20(I)	250(C)	250(I)
Boiler Load Factor (%)	42	55	31	45	40	45	28	36	19	41
Flue Gas Flow Rate (scfm)	6,000	6,000	65,600	65,600	500	500	4,400	4,400	63,300	63,300
Total Capital Requirement for Boiler Modification (10^3 \$)[a]	24.5	24.5	123	123	5.95	5.95	11.0	11.0	70.0	70.0
Annual Operating Cost (10^3 \$/yr)										
Fuel Cost[b]	138.5	181.4	1,278	1,855	13.19	14.84	92.3	118.7	783	1,690
Less Base Case Fuel Cost	-36.1	-47.2	- 333	-483	-4.12	-4.64	-28.9	-37.1	-245	- 528
Investment-Related Cost[c]	3.6	3.6	18	18	0.87	0.87	1.6	1.6	10	10
Effect on Boiler Operating Cost[d]	-14.0	-18.3	- 129	-187	-2.73	-3.07	-19.1	-24.6	-162	- 350
Total	92.0	119.5	834	1,203	7.21	8.00	45.9	58.6	386	822
Sulfur Removed (10^6 lb/yr)[e]	0.333	0.436	3.07	4.46	0.032	0.036	0.222	0.285	1.88	4.06
Control Cost (\$/lb S Removed)	0.276	0.274	0.272	0.270	0.227	0.224	0.207	0.206	0.205	0.202
Control Cost (\$/10^6 Btu Output)	1.25	1.24	1.23	1.22	1.03	1.01	0.94	0.93	0.93	0.92

(a) Based on R. Schreiber, et al., EPA-650/2-74-123, November, 1974, Section 6.

(b) \$1.60/10^6 Btu.

(c) 14.55 percent of investment per year, based on utility financing method with no maintenance.

(d) Based on average 1973 difference in operating cost (excluding fuel) between gas-fired and coal-fired boilers (-0.19 mills/kWh) or between gas-fired and oil-fired boilers (-0.39 mills/kWh). Values from Electrical World, Nov. 1, 1973.

(e) Difference between sulfur emission for base case and sulfur emission at coal gasification plant.

(f) C and I indicate commercial and industrial boilers, respectively.

TABLE 30. CONTROL COST ANALYSIS FOR SOLVENT REFINED COAL (SOLID)

Item	Coal-Fired Boiler, 10^3 lb/hr Steam[e]				Oil-Fired Boiler, 10^3 lb/hr Steam					
	20(C)	20(I)	250(C)	250(I)	2(C)	2(I)	20(C)	20(I)	250(C)	250(I)
Boiler Load Factor (%)	42	55	31	45	40	45	28	36	19	41
Flue Gas Flow Rate (scfm)	6,000	6,000	65,600	65,600	500	500	4,400	4,400	63,300	63,300
Total Capital Requirement for Boiler Modification (10^3 $)[a]	9.9	9.9	48.8	48.8						
Annual Operating Cost (10^3 $/yr)										
Fuel Cost[b]	89.9	117.7	829	1,204						
Less Base Case Fuel Cost	-36.1	-47.2	-333	- 483						
Investment-Related Cost[c]	1.4	1.4	7	7						
Effect on Boiler Operating Cost										
Total	55.2	71.9	503	728						
Sulfur Removed (10^6 lb/yr)[d]	0.189	0.248	1.75	2.53						
Control Cost ($/lb S Removed)	0.292	0.290	0.288	0.287						
Control Cost ($/$10^6$ Btu/Output)	0.750	0.746	0.741	0.739						

(a) Based on R. Schreiber, et al., EPA-650/2-74-123, November, 1974, pp 6-22 & 23.

(b) $1.04/$10^6$ Btu, based on average of four analyses for 7,236 ton/day plants.

(c) 14.55% of investment per year, based on utility financing method with nomaintenance.

(d) Based on solvent refined coal with 15,900 Btu/lb heating value and 0.5 weight percent sulfur.

(e) C and I indicate commercial and industrial boilers, respectively.

TABLE 31. CONTROL COST ANALYSIS FOR SOLVENT REFINED COAL (LIQUID)

Item	Coal-Fired Boiler, 10^3 lb/hr Steam[(f)]				Oil-Fired Boiler, 10^3 lb/hr Steam					
	20(C)	20(I)	250(C)	250(I)	2(C)	2(I)	20(C)	20(I)	250(C)	250(I)
Boiler Load Factor (%)	42	55	31	45	40	45	28	36	19	41
Flue Gas Flow Rate (scfm)	6,000	6,000	65,600	65,600	500	500	4,400	4,400	63,300	63,300
Total Capital Requirement for Boiler Modification (10^3 \$)(a)	262	262	1,477	1,477	127	127	275	275	1,139	1,139
Annual Operating Cost (10^3 \$/yr)										
Fuel Cost(b)	89.9	117.7	829	1,204	8.56	9.63	59.9	77.1	508	1,097
Less Base Case Fuel Cost	-36.1	-47.2	-333	- 483	-4.12	-4.64	-28.9	-37.1	-245	- 528
Investment-Related Cost(c)	38.1	38.1	215	215	18.48	18.46	40.0	40.0	166	166
Effect on Boiler Operating Cost(d)	15.8	20.7	146	212	0.11	0.12	0.7	1.0	6	14
Total	107.7	129.3	857	1,148	23.03	23.59	71.7	81.0	435	749
Sulfur Removed (10^6 lb/yr)(e)	0.189	0.248	1.75	2.53	0.0085	0.0095	0.0593	0.0762	0.503	1.084
Control Cost (\$/lb S Removed)	0.57	0.52	0.49	0.45	2.72	2.48	1.21	1.06	0.86	0.69
Control Cost (\$/10^6 Btu Output)	1.46	1.34	1.26	1.16	3.29	2.99	1.46	1.28	1.05	0.83

(a) Based on R. Schreiber, et al., EPA-650/2-74-123, November, 1974, Section 6.

(b) \$1.04/10^6 Btu, based on average of four analyses for 7,236 ton/day plants.

(c) 14.55 percent of investment per year, based on utility financing method with no maintenance.

(d) Includes cost of additional SRC to heat up and melt the SRC (199 Btu/lb) plus for boilers originally coal-fired the average 1973 difference in operating cost (excluding fuel) between oil-fired and coal-fired boilers (0.20 mills/kWh from Electrical World, Nov. 1, 1973).

(e) Based on SRC with 15,900 Btu/lb heating value and 0.5 weight percent sulfur.

(f) C and I indicate commercial and industrial boilers, respectively.

TABLE 32. CONTROL COST ANALYSIS FOR H-COAL LIQUEFACTION

Item	Coal-Fired Boiler, 10^3 lb/hr Steam[f]				Oil-Fired Boiler, 10^3 lb/hr Steam					
	20(C)	20(I)	250(C)	250(I)	2(C)	2(I)	20(C)	20(I)	250(C)	250(I)
Boiler Load Factor (%)	42	55	31	45	40	45	28	36	19	41
Flue Gas Flow Rate (scfm)	6,000	6,000	65,600	65,600	500	500	4,400	4,400	63,300	63,300
Total Capital Requirement for Boiler Modification (10^3 $)[a]	52.4	52.4	295	295						
Annual Operating Cost (10^3 E/yr)										
Fuel Cost[b]	119.7	156.7	1,105	1,603	11.40	12.83	79.8	102.6	677	1,461
Less Base Case Fuel Cost	-36.1	-47.2	-333	- 483	- 4.12	- 4.64	-28.9	- 37.1	-245	- 528
Investment-Related Cost[c]	7.6	7.6	43	43						
Effect on Boiler Operating Cost[d]	14.7	19.3	136	197						
Total	105.9	136.4	951	1,360	7.28	8.19	50.9	65.5	432	933
Sulfur Removed (10^6 lb/yr)[e]	0.207	0.271	1.91	2.77	0.0101	0.0114	0.0710	0.0913	0.602	1.299
Control Cost ($/lb S Removed)	0.512	0.503	0.498	0.491	← 0.718 →					
Control Cost ($/$10^6$ Btu Output)	1.44	1.42	1.40	1.38	← 0.883 →					

(a) Based on R. Schreiber, et al., EPA-650/2-74-123, November, 1974, Section 6.

(b) $1.38/$10^6$ Btu, based on analysis of plant producing 57,466 bbl/D synthetic crude oil.

(c) 14.55 percent of investment per year, based on utility financing method with no maintenance.

(d) Based on average 1973 difference in operating cost (excluding fuel) between oil-fired and coal-fired boilers (0.20 mills/kWh from Electrical World, Nov. 1, 1973).

(e) Based on feedstock with 18,000 Btu/lb heating value and 0.2 weight percent sulfur.

(f) C and I indicate commercial and industrial boilers, respectively.

employed:

(1) The existing coal-fired boiler system has been equipped with a dust collecting system.

(2) Elemental sulfur and sulfuric acid were assumed to be the by-products resulting from the Wellman-Lord and MgO processes, respectively.

(3) The sludge generated from the FGD processes was assumed to be filtered and disposed of in landfills.

(4) Flue gases from FGD processes would be reheated if necessary using an indirect steam reheat system.

(5) The retrofit factor was assumed to be 1.2 for the double alkali system and 1.3 for the other processes.

(6) The costs for oil-fired boilers were estimated from those for coal-fired with adjustments made with respect to flue gas flow rate and sulfur input.

The summary of the costs of the FGD processes along with other alternatives is shown in Tables 33 and 34 for coal- and oil-fired boilers, respectively.

Physical cleaning of coal appears very attractive in its economics; however, its application is limited to certain types of coal.

Processed fuels produced on a large scale are economically favored over application of FGD processes to boilers in the small size classes. The capital requirement to the boiler system is low and the annualized cost also is low mainly due to the small fixed capital charges.

The FGD processes are economically favorable over other alternatives for the NUC sources in the large size classes. Among the various FGD processes, the MgO process with regeneration performed at a central station appears most attractive economically. Throwaway processes in general are more attractive than integrated regenerable processes if land is available for sludge disposal.

TABLE 33. CAPITAL REQUIREMENT AND ANNUALIZED CONTROL COST OF ALTERNATIVES FOR COAL-FIRED BOILERS

C: Commercial Boiler I: Industrial Boiler

Alternative	Capital Requirement, $10^3				Annualized Control Cost							
					Steam Output, $/10^6 Btu				Sulfur Removal, $/lb S			
	20 (C)	20 (I)	250 (C)	250 (I)	20 (C)	20 (I)	250 (C)	250(I)	20 (C)	20 (I)	250 (C)	250 (I)
Physical cleaning of coal	0	0	0	0	← 0.15 →				← 0.14 →			
Coal gasification (low Btu)	29.7	29.7	153	153	0.77	0.75	0.74	0.73	0.29	0.28	0.28	0.27
Coal gasification (high Btu)	24.5	24.5	123	123	1.25	1.24	1.23	1.22	0.28	0.27	0.27	0.27
Solvent refined coal (solid)	9.9	9.9	48.8	48.8	0.75	0.75	0.74	0.74	0.29	0.29	0.29	0.29
Solvent refined coal (liquid)	262	262	1,477	1,477	.46	1.34	1.26	1.16	0.57	0.52	0.49	0.45
H-coal	32.4	32.4	295	295	1.44	1.42	1.40	1.38	0.51	0.50	0.50	0.49
Fluidized-bed combustion (coal)	443	452	1,892	1,970	2.16	1.90	1.09	0.93	0.63	0.56	0.32	0.27
Wet limestone scrubbing (Peabody)	616	624	2,588	2,625	2.04	1.71	0.94	0.79	0.87	0.73	0.40	0.34
Lime scrubbing (Bahco)	780	780	3,276	3,276	2.50	2.04	1.15	0.87	1.00	0.82	0.46	0.35
Double alkali (FMC)	697	702	2,918	2,962	2.27	1.86	1.11	0.87	0.86	0.70	0.42	0.33
MgO (integrated)	858	871	3,524	3,579	3.32	2.81	1.24	0.95	1.25	1.06	0.47	0.36
MgO (central regeneration)	482	490	2,007	2,028	1.69	1.41	0.69	0.52	0.64	0.53	0.26	0.20
Wellman-Lord	757	765	3,548	3,589	2.76	2.29	1.21	0.92	1.04	0.87	0.46	0.35

TABLE 34. CAPITAL REQUIREMENT AND ANNUALIZED CONTROL COST OF ALTERNATIVES FOR OIL-FIRED BOILERS

C: Commercial Boiler I: Industrial Boiler

Alternative	Capital Requirement, $10^3						Annualized Control Cost											
							By Steam Output, $/10^6 Btu						By Sulfur Removal, $/lb S					
	2(C)	2(I)	20(C)	20(I)	250(C)	250(I)	2(C)	2(I)	20(C)	20(I)	250(C)	250(I)	2(C)	2(I)	20(C)	20(I)	250(C)	250(I)
Coal gasification (high Btu)	6.0	6.0	11	11	70	70	1.03	1.01	0.94	0.93	0.93	0.92	0.23	0.22	0.21	0.21	0.21	0.2
Solvent refined coal (liquid)	127	127	275	275	1,139	1,139	3.29	2.99	1.46	1.28	1.05	0.83	2.72	2.48	1.21	1.06	0.86	0.6
N-coal	← 0 →						← 0.88 →						← 0.72 →					
Wet limestone scrubbing (Peabody)	139	139	498	503	2,456	2,497	5.32	4.91	2.26	1.89	1.29	0.69	4.40	4.06	1.87	1.56	1.06	0.57
Lime scrubbing (Bahco)	173	173	637	637	3,156	3,156	7.53	6.89	2.85	2.30	1.60	0.84	5.84	5.34	2.21	1.78	1.24	1.72
Double alkali (FMC)	154	154	538	540	2,662	2,708	6.85	6.34	2.47	2.01	1.42	0.77	5.04	4.47	1.82	1.48	1.04	0.57
MgO (integrated)			607	612	2,948	3,021			3.30	2.79	1.60	0.89			2.43	2.05	1.17	0.65
MgO (central regeneration)	111	111	393	398	1,925	1,956	4.78	4.46	1.90	1.60	1.00	0.53	3.52	3.29	1.40	1.18	0.73	0.39
Wellman-Lord			602	607	2,929	2,986			3.08	2.58	1.56	0.86			2.27	1.90	1.14	0.63

PART III

PACKAGEABILITY OF SORPTION PROCESSES

SURVEY OF EXISTING PACKAGE SORPTION SYSTEMS

A "package system" is loosely defined here as a complete, compact, factory assembled, and easily transportable unit or combination of units, and/or a unit which can be assembled on site from a few prefabricated and readily transportable components. Such systems are characterized by: (1) small size, (2) complete integral components, (3) simple and easy installation, (4) ready availability, (5) ease of operation and maintenance, and (6) low capital and operating costs.

Package systems in general have been widely used in many areas such as household appliances, wastewater treatment plants for small establishments, and solvent recovery and air purification from exhaust and vent gases.

In order to obtain information on existing package sorption systems, if any, applicable to removal of SO_x from small commercial and industrial boilers, a literature search was conducted extensively using information systems such as National Technical Information Service (NTIS), Stack Gas Control Coordination Center (an organization within Battelle-Columbus), Battelle's Energy Information Center base, Engineering Index System, Atomic Energy Commission (AEC) system, and Battelle-Columbus libraries. In addition, manufacturers of sorbent materials and engineering and fabrication companies producing sorption systems were contacted to acquire the field information on existing package sorption systems for SO_x control.

The survey results indicate that for the prupose of recovering solvents and purifying circulating air, activated carbon is used extensively mainly because it adsorbs all types of organic vapors and mists regardless of variation in concentration and humidity. Factory assembled solvent recovery systems that are essentially off-the-shelf units are commercially available up to 10,000 scfm flow capacity. Three mechanically different types of systems, fixed bed, moving bed, and fluidized bed, are being utilized. However, as Tables 35 and 36 indicate, currently there is no package sorption system commercially available for SO_x control. This is attributed to the fact that the marketability of such a system is low since industry or a potential buyer would not invest its capital where there is no return on investment or where there is no OSHA requirement. In addition, according to the opinion

TABLE 35. SURVEY OF ENGINEERING AND FABRICATION COMPANIES OF SORPTION SYSTEMS

Company	Sorption Systems	Package Sorption System for SO_x	Remarks
American Air Filter Company, Inc.	Lime and limestone slurry	NA	Custom design only
Day and Zimmerman, Inc.	Carbon resorb system	NA	Custom design only
Mine Safety Appliances Company	Carbon based air purification	NA	Panel and pleated carbon beds
T. Malsheimer Company, Inc.	Multistage beds for air purification	NA	California Carbon Company provides regeneration of carbon
Vulcan Cincinnati, Inc.	Solvent recovery, stationary carbon bed	NA	Custom design
Cambridge Filter Corp.	Carbon based air purification	NA	
Vic Manufacturing Company	Solvent recovery	NA	Custom design; small packaged system available
Farr Company	Carbon based air purification	NA	Regeneration at California Carbon Company
Howard S. Caldwell Company	Carbon based air purification	NA	
Detrex Chemical Industries, Inc.	Solvent recovery	NA	Custom design of compact system

TABLE 35. (Continued)

Company	Sorption Systems	Package Sorption System for SO_x	Remarks
RSE, Inc.	Carbon panel air purification	NA	Package system for air purification
Conner Engineering Corp.	Carbon panel and radial flow canister air purification	NA	Standard system available
Hoyt Manufacturing Corp.	Solvent recovery	NA	Cabinet enclosed system

TABLE 36. SURVEY OF SORBENT MANUFACTURERS

Company	Activated Carbon	Noncarbon Sorbent	Package Sorption System for SO_x	Remarks
Borg-Warner Corp.		Potassium permangated impregnated alumina	NA[(a)]	Sold the plant to H. E. Burroughs & Assoc.
Barnebey-Cheney Company	X		NA	Granular carbon
Sude Chemie U.S. Company	X	Zinc oxide catalyst	NA	Catalyst for H_2S sorption
C. H. Dexter Corp.	X		NA	No interest in SO_x control
American Norit Company, Inc.	X		NA	
Witco Chemical Company, Inc.	X		NA	Activated carbon for air conditioning equipment
Aluminum Company of America		Granular activated alumina	NA	
Husky Industries	X		NA	Powdered carbon for water treatment
Davison Chemical		Silica gel Molecular sieve	NA	
ICI United States	X		NA	Carbon for liquid phase adsorption
Westvaco	X		NA	Regenerable FGD process
Union Carbide Corp. at Fostoria, Ohio	X		NA	
Linde Company		Molecular sieve	NA	Purasiv S system for sulfuric acid plant
Calgon Corp.	X		NA	SO_2 control process similar to Sulfacid

(a) NA - not available.

survey, activated carbon, which has been used extensively in package sorption systems, would not be adequate for SO_x control since its sorption capacity for SO_x is relatively small (i.e., generally less than 10 percent by weight) and the regeneration requires a special feature for handling sulfurous regeneration products. The central regeneration mode of the operation also requires the transportation of bulky activated carbon, and thus, a dry activated carbon process is not deemed feasible for a package system to control SO_x.

SURVEY OF SORBENT MATERIALS

The potential package sorption systems were surveyed by reviewing the applicability of sorbent materials to the removal of SO_2 from boiler flue gases and examining the current development of various flue gas desulfurization (FGD) processes using the sorbent materials.

Sorbent materials which can be used to control SO_x may be classified into solid sorbent, aqueous solution, and organic liquid. The description of each of the sorbent materials follows.

Solid Sorbents

A number of solid materials react with sulfur dioxide under suitable conditions and can be utilized to remove it from flue gases. The most important requirement for a solid sorbent, in addition to a high affinity for sulfur dioxide, is that it has a large surface area. The area can be increased by either granulating the solid or making it very porous. In granular form the sorbent can be contacted by the flue gas as a fluidized or gravitating bed or as entrained particles. The gas may contact the interior of porous solids by diffusion through the pores. Nonporous sorbents may be modified to give increased exposure to the gas by impregnating them on more porous inert materials.

Most solid sorbent processes are dry and have the advantage that the flue gas is not cooled. Unlike aqueous solutions, little, if any, solid or liquid wastes are generated, eliminating the problem of waste disposal.

With the exception of limestone, most solid sorbents are expensive and must be regenerated. In addition, solid sorbents tend to lose their activity due to contamination and plugging of the sorbent pores by solid impurities in the gas stream.

Dry Limestone. In the dry state and at ordinary temperatures, neither limestone ($CaCO_3$) nor lime (CaO) react well with SO_2.(64) At high temperatures, however, on the order of 1800°F, the $CaCO_3$ is quickly calcined to CaO and readily reacts with the SO_2.

Limestone is very abundant and inexpensive. There are, however, intrinsic disadvantages. That is, the low porosity of the calcined limestone particles and the narrowing of the exterior pores due to the formation of $CaSO_4$ result in a low degree of SO_2 adsorption and consequently a low removal efficiency.

Activated Carbon. Activated carbon may be utilized as a sorbent for low temperature (less than 300°F) adsorption of SO_2 from flue gases.(65-67) The SO_2 diffuses into the pores of the carbon matrix and is catalytically oxidized to SO_3 by contacting active carbon sites. The presence of water in the flue gas converts the SO_3 to sulfuric acid freeing the active sites for further adsorption.

The acid may be removed from the carbon structure by washing with water or alkaline solution, by thermally desorbing with a hot scavenger gas, or by reducing with a reducing gas such as H_2S. Water washing will produce a stream of weak sulfuric acid, while thermal desorption, conducted at 750°F, utilizes the carbon as a reductant to generate a concentrated SO_2 off gas. Reduction with H_2S will generate elemental sulfur.

Carbon adsorbent may be manufactured from a variety of carbonaceous materials. Charcoal and semicokes prepared from coal, lignite, and peat are suitable adsorbents. Lignite is more applicable to a dry process employing thermal regeneration. Upon successive adsorption-regeneration cycles, the porosity and activity of the lignite tend to increase.

Two primary drawbacks to all carbon-based systems are the limited capacity and low gas velocity requirements. The sorbent capacity is generally only about 2-10 percent sulfur by weight. This necessitates the use of large quantities of carbon adsorbent. In addition, the rate of adsorption of SO_2 on carbon is limited to the rate of SO_2 diffusion into the pores. Thus, to allow time for the diffusion the gas velocities must be limited to 1-4 feet per second. Carbon, however, has several characteristics which make it an attractive sorbent for SO_2.

(1) Carbon is easily regenerated and is amenable to a process utilizing separate independent adsorption and regeneration options.

(2) Carbon-based processes are generally simple.

(3) The regeneration operations are not complicated by side reactions.

(4) In dry carbon processes, adsorption may be carried out at air preheater temperatures (300°F), thus no flue gas reheat is required.

(5) No waste streams are generated.

(6) In wet carbon processes preceded by adequate particle removal equipment, the sorbent has a long, almost indefinite life.

Metal Oxides. The oxides of Mn, Fe, Cu, and Zn will react strongly with SO_2 at elevated temperatures and have been investigated for utilization as sorbents to remove SO_2 from flue gases.(68-70) They have several characteristics that make them applicable to flue gas desulfurization systems.

(1) High affinity for SO_2.

(2) Adsorption process is dry, at air heater temperature of 700°F, eliminating need for flue gas reheat.

(3) No generation of waste streams.

(4) Regenerable.

Of the above sorbents, the Mn and Cu oxides have been found to be the most promising. According to the studies done by the United States Bureau of Mines and the TVA, MnO was superior to other oxides of metals for absorbing SO_2. The process, however, involves a complicated regeneration process.

Cupric oxide is presently the only metal oxide sorbent being actively tested in the United States. Sulfur dioxide and oxygen, upon contacting the cupric oxide, react rapidly with it to form $CuSO_4$. The optimum operating temperature is about 700°F. Cupric oxide granules do not have sufficient porosity to function as an effective sorbent. To increase the porosity, they are impregnated on an activated alumina acceptor. The optimum copper content of the acceptor appears to be 4-6 percent by weight.

The sorbent/acceptor is expensive and must be regenerated. Moreover, several problems, intrinsic to the sorbent, dictate the design and operating conditions of the absorption-regeneration system. At high temperatures or under thermal stress, the activated alumina is attacked by SO_3, forming $Al_2(SO_4)_3$, which stresses and weakens the alumina structure. To minimize the thermal stress, the absorption and regeneration reactions must operate at similar temperatures. This is why thermal regeneration, which requires a temperature of 1300°F, would not be acceptable. Regeneration by reduction, however, requires a much lower temperature. The $CuSO_4$ may be contacted by hydrogen or methane at 750°F to regenerate the copper and drive off a concentrated stream of SO_2.

Physical movement or transport of the acceptor results in degradation and disintegration of the matrix structure. Moving bed or continuous flow systems would not be acceptable as they would subject the sorbent to excessive movement and vibration. A fixed bed system would be required.

Tests conducted on a Shell pilot plant using reduction regeneration and fixed, parallel beds indicated an acceptor life of about 1.5 years.

The pore structure is easily contaminated and plugged by particulate matter. To help alleviate this problem the absorber must be preceded by a highly efficient electrostatic precipitator.

Another problem associated with the regenerable cupric oxide absorbent is the production of sulfur dioxide gas. It is an intermediate by-product and requires processing to convert it to a more marketable form such as sulfur, sulfuric acid, or liquid SO_2.

Oxidation Catalyst. The catalytic oxidation process for sulfur dioxide removal from flue gases is based on the same concept as the contact process for manufacturing sulfuric acid.(71) The best known catalyst is vanadium pentoxide impregnated on an alumina support. At the optimum operating temperature of 800-900°F, about 90 percent of the SO_2 is converted to SO_3 on the catalyst. Following oxidation the gas is passed through an absorption tower where the SO_3 is removed by contacting a countercurrent solution of sulfuric acid.

The vanadium pentoxide based process has some unique advantages over sorption-regeneration processes.(72)

(1) It is a simple process requiring only two steps. No regeneration is involved.

(2) No waste side streams are produced.

(3) Except for catalyst replacement, no raw materials are consumed.

(4) The temperature of the gas leaving the absorber is 250°F; no flue gas reheat is necessary.

The catalytic oxidation process, however, is accompanied by some operational difficulties.

(1) The catalyst is easily plugged by particulate matter and requires frequent replacement. The catalyst is expensive and removal from the converter for cleaning generally results in attritional losses.

(2) The catalytic converter must be operated at a high temperature. A retrofit process would require flue gas preheat facilities and heat exchangers to minimize energy losses.

Molecular Sieve. Molecular sieves can be utilized as effective adsorbents for the removal of SO_2 from flue gases. The molecular sieve most applicable to SO_2 removal is a crystalline metal aluminosilicate. It is highly porous and selectively adsorbs polar molecules like SO_2 and H_2O.(73)

Sulfur dioxide removal with molecular sieves is a relatively simple process, requiring two steps, adsorption and regeneration. The removal efficiency is high; for example, in sufficient amounts the molecular sieve can remove greater than 99 percent of the SO_2 from the flue gas.(73)

However, problems, intrinsic to the nature of the sorbent, eliminate it as a process applicable for boiler flue gas treatment.

(1) The molecular sieve adsorbs water vapor with the SO_2. Since water vapor in boiler flue gas streams is considerably greater in concentration than the

SO_2, the sieve would be quickly contaminated or loaded with water. An additional sieve would have to be installed ahead of the SO_2 adsorber to selectively remove the water vapor.

(2) The sorption capacity of a molecular sieve for SO_2 is decreased substantially with increases in temperature. The loading can vary from 6 percent SO_2 by weight at 105°F to 13 percent at 50°F. For this reason, it is mandatory that the adsorption temperature be maintained as low as possible. This is difficult for boiler flue gas streams which are generally at about 300°F. An acid resistant heat exchanger would have to precede the adsorber.

(3) Other problems associated with the process are the necessity for flue gas reheat and installation of facilities to process the SO_2 off gases.

Alkalized Alumina. Alkalized alumina ($NaAlO_2$) is an effective absorbent for SO_2. At 570°F to 660°F the sulfur dioxide readily reacts with the alkalized alumina to form sodium sulfate and Al_2O_3. The alumina is included to provide increased porosity for the sodium salt and to assist in the regeneration process. During regeneration the alumina provides an anion to combine with the Na^+ and reduces the reaction energy requirements. Regeneration is accomplished by reduction with H_2 gas at 1350°F. The H_2S off-gas can be utilized in a Claus reactor for the production of elemental sulfur.

Alkalized alumina has several characteristics which make it an attractive sorbent for SO_2 removal systems.

(1) The absorption kinetics favor reaction with the SO_2 at near flue gas preheater effluent temperatures. Retrofit is easy and no flue gas is necessary.

(2) It has a high affinity for SO_2. Ninety percent removal efficiencies are attainable with a reasonable amount of absorbent.

(3) The sorption capacity is high, about 15-20 percent sulfur dioxide by weight.

(4) There are no adverse side reactions or waste streams.

The sorbent does, however, have some drawbacks. It is very costly and requires regenerating. The fragile crystal structure degrades rapidly from the mechanical recycling operation to and from the absorption and regeneration vessels and the thermal stress created by the high regeneration temperatures. Operating costs, resulting from the replacement of the disintegrated sorbent, are estimated to be very high.(67, 74)

Another problem is the generation of the H_2S off-gas. It requires processing with SO_2 in a Claus reactor to produce elemental sulfur.

Organic Solids. The application of organic solids processes for the removal of sulfur dioxide has received relatively little attention as compared to other processes. This is primarily due to the inherent properties of most solids, i.e., low adsorption capacity and adsorption rates, poor thermal stability and high cost.

Laboratory studies were conducted on the suitability of organic ion exchange resins for sulfur dioxide removal.(75) However, the adsorption rates were slow and could not compare with that of molecular sieves or carbon. Other studies indicated that successive adsorption-thermal regeneration cycles lead to the decrease of the adsorption capacity of the resin.(76)

A study examining nitrogen containing polymers, incorporated into melt-spun fibers, for use as adsorbents revealed that styrene-dimethylpropylmalimide was the most applicable.(77) It was subject to a build up of sulfate, however, and its adsorption capacity decreased with each adsorption-regeneration cycle.

An investigation by TRW on several organic solids showed that waste newsprint can adsorb up to 10 percent of its weight in sulfur dioxide. The adsorption rate and resultant flow velocities were very slow however.(78)

Aqueous Solutions

Aqueous solutions have received the most attention for removing SO_2 from flue gases.(79, 80) They may be classified into three different categories: slurry solutions, clear solutions, and weak acid solutions.

Slurry Solutions. Slurry solutions involve the use of a 5-15 percent lime or limestone slurry to absorb SO_2. The spent absorbent, $CaSO_3$ and $CaSO_4$, is transferred to a disposal facility and discarded as waste.

The attractive feature of the lime/limestone based scrubbing system is that the low cost of the absorbent eliminates the need for regeneration facilities. Alternately, provisions must be made to dispose of the $CaSO_3$ and $CaSO_4$ waste.

The lime or pulverized limestone slurry contacts the SO_2 in the scrubber and reacts to form $CaSO_3$ and $CaSO_4$. Although the reaction products are the same for lime or limestone, lime is a more effective adsorbent than limestone. This may be because the $CaCO_3$ particles formed from lime are smaller than the limestone particles and hence have more surface area and greater reactivity.(67) Moreover, the calcium from lime is already in solution, whereas the calcium from the natural $CaCO_3$ is not, and must undergo dissolution to calcium ion.

Magnesium in the form of dolomitic lime or limestone may be added to either system to improve the absorption efficiency. The higher solubility of $MgSO_3$ permits higher ion concentrations of absorbents in the solution. Pilot plant tests have demonstrated that a higher SO_2 removal and improved operation and maintenance are possible by adding magnesium into the scrubbing solution.(81)

One critical problem with lime/limestone slurry scrubbing is the deposition of solids on surfaces in the scrubber and associated equipment from crystallization of calcium sulfite or calcium sulfate. The solubility of calcium sulfite is very sensitive to pH variations and decreases with increasing pH. The liquor entering the scrubber is saturated with calcium sulfite. As the liquor proceeds through the scrubber, SO_2 is absorbed, the pH decreases, and $CaCO_3$ dissolves and reacts with SO_2 to form $CaSO_3$. The newly formed calcium sulfite either supersaturates or oxidizes to $CaSO_4$. Any calcium sulfite in excess of the saturation equilibrium concentration formed during absorption must be removed in the hold tank prior to recycling to the scrubber. This is accomplished by raising the pH with the addition of lime or limestone. If the slurry recycled to the scrubber contains supersaturated sulfite, the potential for scaling exists.

Calcium sulfate is formed from the oxidation of calcium sulfite. Unlike calcium sulfite, the sulfate is not precipitated from solution by the rise in pH that occurs in the hold tank and, consequently, accumulates until it is supersaturated at which time it crystallizes resulting in scaling of the scrubber internals. Three of the several methods which have been employed to alleviate the supersaturation conditions are: (a) adding seed crystals to the hold tank to maximize crystallization, (b) providing maximum residence time for the slurry in the hold tank to enable the sulfate to precipitate, (c) adding magnesium, as MgO, to form the more soluble $MgSO_4$ and reduce the saturation of the sulfate, and (d) keeping oxidation of sulfite to sulfate at a low level.

An important consideration, with coal fired boilers is the influence of chloride on scrubbing performance. The chloride enters the scrubber with the flue gas as HCl and because it forms no insoluble compounds with calcium, it accumulates in the scrubber slurry. It can only leave the system with the liquid purge, i.e., filter cake moisture or clarifier underflow. Since the liquid purge from a closed system is small, high concentrations, <5000 ppm, of chloride can exist in the slurry.

The addition of chloride lowers the pH of the scrubbing solution. Since oxidation increases with decreasing pH, an increase in chloride ion increases the sulfate content.(81) The chloride ion also increases the calcium ion concentration resulting in decreased dissolution of calcium sulfite. Consequently, the calcium sulfite available for reaction with sulfur dioxide decreases, resulting in reduced scrubbing efficiency and forcing the utilization of higher liquid to gas ratios.

The generation of huge volumes of calcium sulfite and calcium sulfate sludge poses an additional problem. The sludge is of little commercial value and must be disposed of as a waste product. The most environmentally sound methods for sludge disposal are landfilling of chemically fixed sludge and disposal of untreated sludge in ponds lined with an impervious material such as clay, plastic, or rubber. However, calcium sulfite presents a significant land use problem. Sulfites tend to crystallize into small, thin platelets which settle to a loose bulky structure that may occlude a relatively large amount of water. The sludge is difficult to compress and dewater, and conversion to a suitable landfill presents an expensive and formidable problem. Ponds on the hand require a large, suitable disposal site close to the plant and may be not only structurally unstable but aesthetically objectionable.

Other problems associated with wet lime/limestone slurry scrubbing are

(1) Flue gas reheat is necessary

(2) For limestone systems, limestone grinding facilities must be installed or provisions must be made to purchase the limestone in pulverized form

(3) The abrasive slurry solutions may cause pump and equipment erosion problems.

Clear Solutions. Clear scrubbing solutions involve the use of alkali absorbents to remove sulfur dioxide from flue gas. The most common alkali absorbents are sulfites of ammonia, sodium, potassium, and lithium. The more popular scrubbing solutions are ammonium sulfite and sodium sulfite. Ammonia is reasonably priced, permits a variety of regeneration procedures, and in the sulfate form, has commercial value as a fertilizer. Sodium, although higher priced than ammonia per equivalent, is nonvolatile, eliminating the problem of absorbent volatilization in the scrubber. Potassium and lithium solutions have also been used but they are more costly than the ammonia or sodium absorbents.

The sulfur dioxide reacts with alkaline salt in the form of sulfite to produce the bisulfite. In cases where the starting material is a metal oxide or hydroxide, its reaction with sulfur dioxide forms the sulfite first followed by conversion to the bisulfite.

Clear solutions have three characteristics which make them attractive absorbents for SO_2.

(1) They have a high affinity for SO_2 and are generally capable of greater than a 90 percent removal efficiency.

(2) All of the compounds are highly soluble and remain in solution minimizing plugging and scaling problems associated with other wet processes like lime or limestone.

(3) The sorption capacity is high, resulting in the use of less amounts of sorbent in the operation.

Alkali absorbents are not inexpensive and thus should be either regenerated or transformed into a marketable by-product. Some methods generally employed to process the spent absorbent are:(67)

(1) Direct thermal treatment - The sulfite-bisulfite chemistry is applicable to an adsorption-thermal regeneration type system. The bisulfite is relatively unstable and SO_2 can be desorbed from it relatively easily. A problem with thermal regeneration is the large amount of heat consumed. Most clear solutions contain less than 20 percent salt by weight and consequently the heat supplied must be expended in evaporation of water.

(2) Acidification - This process is applicable to ammonia scrubbing systems. Acidulation of the ammonium bisulfite and sulfite with H_2SO_4, HNO_3, or H_3PO_4 will produce SO_2 and either ammonium sulfate, ammonium nitrate, or ammonium phosphate as a fertilizer by-product.

(3) Reduction - Direct reduction of the sulfite with H_2S will regenerate the absorbent and form elemental sulfur. The United States Bureau of Mines Citrate process bubbles H_2S through the spent absorbent followed by separation of the by-product sulfur and recovery of the citrate solution for recycle to the scrubber.(82)

(4) Oxidation - This process also is applicable to ammonia systems for generating a marketable by-product. Air may be bubbled through a spent ammonium sulfite, ammonium bisulfite solution to oxidize it to ammonium sulfate. The product may be evaporated, dried, and marketed as fertilizer.

(5) Lime or limestone regeneration - This process has been applied to sulfite buffer systems where lime or limestone is added to the spent sodium sulfite, sodium bisulfite solution to regenerate the sodium sulfite and remove the SO_2 as insoluble $CaSO_3$. The FMC and GM double alkali processes utilize this method.(83, 84)

There are some characteristic disadvantages of clear solutions:

(1) The sulfite tends to oxidize to the sulfate. The sulfate is very stable, unreactive, and not easily reconverted to the sulfite.

(2) The processes using clear solutions are wet processes and cool the flue gas down to 120 to 140°F. To increase buoyancy and eliminate the plume, a flue gas reheat system is necessary.

(3) The thermal and acidification-based regeneration systems require an accompanying sulfur dioxide processing facility.

Weak Acid Solutions. An aqueous process not involving the use of alkali salts or calcium compounds utilizes a ferric sulfate catalyst in solution to oxidize SO_2 absorbed as H_2SO_3 to H_2SO_4 such as in the Chiyoda process. The weak acid can be neutralized with limestone to form insoluble gypsum.

The use of the weak sulfuric acid as a sorbent offers several advantages:(85)

(1) It lends itself to an easy to operate, uncomplicated system. Only three processes are involved: absorption, oxidation, and neutralization.

(2) The ferric sulfate catalyst concentrations are weak, only 2000 ppm, and the loss with the gypsum is negligible.

(3) The sorbent will not plug, scale, or erode the equipment.

However, the absorption capacity of water for sulfur dioxide is quite low. To effectively absorb the sulfur dioxide, very high liquid flow rates must be used. Normal liquid to gas ratios (L/G) are on the order of 300. This implies that a very large absorber is required. Furthermore, the sorbent is highly corrosive so all process equipment must be constructed of expensive stainless steel.

Organic Liquids

Organic liquids have found only very limited application as sulfur dioxide sorbents. Although most amines, alcohols, amides, imides, ketones, and esters are capable of absorbing sulfur dioxide, only two, dimethylaniline and xylidene, have been used on limited industrial scale. Both applications are on smelter gases containing about 4-6 percent sulfur dioxide.(86)

One problem in common to most liquid organic sorbents is the loss of the purified gas stream. Although the sorbent may have a low vapor pressure, significant losses can still occur if a large volume of gas is treated. Organic liquids are costly, and thus, the loss can contribute substantially to the operating costs. Furthermore, the loss of the organic sorbent to the atmosphere can pose a potential pollution threat.

DESCRIPTION OF SORPTION PROCESSES

The candidate sorption processes evaluated with respect to the possibility of being a package unit for small industrial and commercial boilers included

(1) Throwaway Process

Limestone slurry (Peabody)
Lime scrubbing (Bahco)
Double alkali (GM and FMC)

(2) Regenerable Process (central regeneration)

MgO process - central regeneration

(3) Regenerable Process (on-site regeneration)

Chemiebau
Foster Wheeler
Westavco
Sulfacid
Chiyoda
Ammonia process (Peabody and Catalytic)
MgO process - integrated
Wellman-Lord
Shell FGD
Citrate (Morrison-Knudsen)
Calsox (Monsanto)
Aqueous carbonate (Atomic International)

The description of the following processes was not included in this section since they have been included in Part II of this report.

Limestone slurry (Peabody)
Lime scrubbing (Bahco)
Double alkali (FMC)
MgO process (central regeneration)
MgO process (integrated)
Wellman-Lord

Below are brief process descriptions; more complete descriptions are included in Appendix B.

Double Alkali Process (GM)

General Motors Corporation (GM) has developed one of a number of double alkali processes that scrub with a Na_2SO_3 buffer solution and then react the clear solution with lime or limestone to precipitate $CaSO_3$.(87) Like FMC Double Alkali processes, the purpose of separating scrubbing from precipitation is to eliminate scaling difficulties. Unlike the FMC process, however, the GM process employs a weaker solution and thus oxidation may be more significant than in the FMC process. The removal efficiencies of SO_2 and particulate matter are about 90 percent and 83 percent, respectively. A full-scale facility on a combined coal-fired combustion source equivalent to 400,000 lb steam/hr capacity in Parma, Ohio, has been in operation since March, 1974.

Chemiebau Process

Reinluft GmbH, Essen, Germany, developed this process and Commonwealth Associates, Jackson, Michigan, has acquired the Western hemisphere licensing rights. The process employs moving beds of lump char to remove SO_2 from flue gas. The deactivated adsorbent is thermally regenerated, producing a concentrated 20 percent SO_2 gas stream.(88, 89) Since char is lost due to mechanical attrition and reaction with SO_3, the makeup char must be added to the system at a rate of 20 lb/lb of SO_2 removed to replace these losses. A heat requirement of 5,000 to 6,000 Btu/lb of SO_2 removed has been estimated for the thermal regeneration of the deactivated char. The removal efficiency of SO_2 ranges from 85 to 95 percent. To date no Chemiebau process has been sold in the United States.

Foster Wheeler (FW)

The FW process for SO_2 removal is a combination of char adsorption and regeneration processes developed by Bergbau Forschung, GmbH and elemental sulfur conversion process developed by Foster Wheeler, who currently markets the process in the United States. After treatment for removal of particulate matter, the flue gas is introduced into the adsorber which contains vertical parallel louver beds through which the char flows.(90, 91) The flue gas passes through the adsorber bed in a cross flow. The deactivated char is regenerated thermally by mixing with hot sand in a fluidized bed and SO_2 gas is liberated. The concentrated SO_2 gas stream is directed to the Foster

Wheeler off-gas treatment system where SO_2 reacts with crushed coal and is reduced to elemental sulfur. The process is capable of removing 86-95 percent of the SO_2, 90-95 percent of the particulate matter, and 40-60 percent of the NO_x. About 0.14 lb of char/lb of SO_2 removed is required to make up the losses primarily due to the production of CO_2 in the regeneration. Fuel is required to heat the sand to 1500°F. About 0.5 lb of elemental sulfur is produced per pound of SO_2 removed. A test program is underway for a demonstration unit on a 47.5 MW coal-fired boiler at Gulf Power Company's Scholz Steam Plant.

Westavco Process

Westavco, Charleston Heights, South Carolina, has developed and markets the process. After treatment for removal of particulate matter, the flue gas is introduced to an activated carbon fluidized bed where SO_2 is removed through catalized oxidation to SO_3 and a subsequent hydrolysis to sulfuric acid which remains adsorbed in the carbon particle. The acid loaded carbon is contacted by a stream of H_2S which reduces the sulfuric acid to elemental sulfur.(92, 93) The system is capable of removing 90 percent SO_2 from flue gas. Some carbon is lost due to mechanical attrition, normally less than 1 percent per cycle. Fuel oil is consumed in the H_2S generator/sulfur stripper to raise the temperature up to 1200°F. Hydrogen must be produced from a coal gasifier to generate H_2S. Westavco has recently completed pilot plant tests on a 20,000 ft^3/hr flue gas stream from an oil-fired boiler and they are interested in evaluating the system on a coal-fired boiler in an increased scale, i.e., 15 MW.

Sulfacid Process

Lurgi of Frankfurt, West Germany, has developed the process and the Rust Engineering Company, Birmingham, Alabama, markets the system in the United States. Stack gas is pretreated to adjust the temperature, humidity, and fly ash content and the conditioned gas flows upward at low velocity through a bed of carbon-based catalyst of 1 or 2 feet deep. SO_2, oxygen, and water are adsorbed on the impregnated carbon where sulfuric acid is formed.(94, 95) The acid is washed from the bed by a continuous spray of water as it is formed. A SO_2 removal efficiency of 90 percent can be achieved by the process. About 10 to 30 percent sulfuric acid is produced as the by-product. If there is no market available for the acid, it should be neutralized with limestone for disposal. The process has been commercially applied to the treatment of chemical plant waste streams for several years in Europe. A plant to handle 20,000 to 30,000 cfm of sulfuric acid tail gas will be built in Pittsburgh, Pennsylvania, for the United States Steel Company.

Chiyoda Process

The Chiyoda Chemical Engineering and Construction Company, Japan, developed the "Thoroughbred 101" sulfur dioxide process. Following treatment by an ESP and a venturi type wet prescrubber the flue gas flows upward through a packed bed absorber and contacts a countercurrent 2 to 5 percent sulfuric acid solution, containing about 2,000 ppm of ferric sulfate.(85, 96) The solution absorbs SO_2 in the flue gas and flows to the oxidizer tower, where, in the presence of the ferric ion catalyst, air injected into the liquor oxidizes the H_2SO_3 to H_2SO_4. A portion of the liquor leaving the oxidizer returns to the absorber and the remainder passes to the gypsum production steps where the liquor is neutralized with lime or pulverized limestone to form insoluble gypsum which then is separated from the mother liquor by centrifuging. The process is capable of a 95 percent SO_2 removal efficiency. Ten commercial plants have been installed on Claus plants and oil-fired boilers in Japan. In the United States, a demonstration program for a 23-MW pilot plant on Gulf Power Company's coal-fired boiler at Scholz plant in Sneads, Florida, is under way.

Ammonia Scrubbing Process (Peabody)

The Peabody Engineering Company, Stamford, Connecticut, has developed the process. After pretreatment to adjust fly ash content, temperature, and humidity, the flue gas is introduced into an abosrber where the gas is scrubbed countercurrently with a mixture of ammonia, ammonium sulfite, ammonium bisulfite, and ammonium sulfate.(97, 98) SO_2 in the gas reacts with ammonia and ammonium sulfite in the solution to form ammonium bisulfite. Spent liquor is pumped to the neutralizer where ammonia is added to convert the bisulfite to sulfite to reduce sulfur dioxide loss during oxidation. The neutralized solution is subsequently treated in an oxidizer where the sulfite is reacted with air to form the sulfate which is pumped to a double-effect vacuum evaporator crystallizer. Greater than a 90 percent removal efficiency of SO_2 can be obtained with the process and ammonium sulfate is the by-product. Although the individual step of the process has been utilized in the ammonia industry, the process has not been tested for flue gases from coal-fired boilers.

Ammonia Scrubbing Process (Catalytic, Inc.)

Catalytic, Inc., a subsidiary of Air Products and Chemicals, Inc., developed the process.(99, 100) The absorption steps of the process are similar to those for Peabody ammonia scrubbing process. However, unlike Peabody ammonia scrubbing process, this process utilizes the Institut Francais du Petrole (IFP) sulfur reducing process to regenerate the spent ammonium salts and produce sulfur. In 1970 IFP began marketing its Claus tail gas treating process which involves ammonia scrubbing coupled with the reduction-regeneration system. To date, seven installations have been constructed and all are currently operating. The complete IFP SO_2 removal system is being installed on a 35 MW utility boiler in France.

Shell Flue Gas Desulfurization Process (Shell FGD)

Shell International Petroleum developed the process in the early 1960's and Universal Oil Products (UOP) purchased the licensing rights for the United States.(101-103) After treatment for removal of fly ash, the flue gas is passed through an adsorber which is a fixed bed of elemental copper supported on an alumina structure with open channels along the side.

The elemental copper reacts with oxygen and SO_2 to form $CuSO_4$. The deactivated bed is arranged for regeneration by passing a stream of hydrogen through the bed. The conversion of $CuSO_4$ to Cu, H_2O, and SO_2 takes place. A cyclic operation can be arranged with two or more identical adsorbers for continuous processing of flue gas. The process can remove 90 percent of the SO_2 from flue gas and hydrogen consumption is about 0.1 lb/lb of SO_2 removed. To date, no commercial plant has been built in the United States.

Citrate Process

The U. S. Bureau of Mines developed the process and the Morrison-Knudsen Company, Inc., Boise, Idaho, and Peabody Engineering Company, Stamford, Connecticut, independently offer the process on a commercial basis.(82) Basically the process involves absorption of SO_2 by a solution of sodium citrate, citric acid, and sodium thiosulfate followed by reacting the absorbed SO_2 with H_2S to precipitate elemental sulfur and regenerate the citrate solution. The major difference of the Morrison-Knudsen process and the Peabody process is the flotation method for elemental sulfur separation. SO_2 removal efficiency ranges from 95 to 99 percent. Small scale pilot plant testing (1000 to 2000 cfm) is under way in Kellogg, Idaho, and Terre Haute, Indiana, by the Morrison-Knudsen and Pfizer-McKee-Peabody, respectively.

Calsox Process

The Monsanto Company of St. Louis, Missouri, developed the process which utilizes a dilute ethanolamine solution (0.5 weight %) to absorb SO_2 from flue gas.(104) Spent liquor is regenerated by using lime and calcium sulfite and calcium sulfate sludges are formed. The process can achieve a 90 percent SO_2 removal efficiency. The loss of high cost ethanolamine is the main disadvantage of the process. A 3000 cfm pilot plant was operated at a boiler owned by the Indianapolis Power and Light Company. To date no large-scale commercial plant is available.

Aqueous Carbonate Process

The Atomic International Division of Rockwell International Corporation developed the process in the 1970's. The flue gas is introduced into the spray tank dryer where it co-currently contacts an atomized mist of a 4 to 20 weight percent Na_2CO_3 solution.(105) The SO_3 in the gas reacts with the sodium carbonate to form sodium sulfite and sodium sulfate. The particulate matter is separated from the flue gas and either disposed of in the open loop operation or regenerated in the closed loop operation. Open loop operation is less costly in a small-scale operation, but the soluble Na_2SO_3 and Na_2SO_4 salts present a disposal problem. In closed loop operation, the scrubbing and regeneration systems are independent and can be uncoupled and operated separately. The regeneration system involves three steps: reduction of sodium sulfite and sulfate to sodium sulfide with either coke or coal, reformation of Na_2CO_3 from the sodium sulfide by dissolving in water and treating with a CO_2-rich gas, and conversion of H_2S resulting from the second step to elemental sulfur by a Claus process. The process can achieve greater than 90 percent SO_2 removal. Although both scrubbing system and the regeneration system have been tested separately, the complete aqueous carbonate process has not been tested in a single installation.

SORPTION PROCESS EVALUATION

Approach

The evaluation of the potential role of the stack gas cleaning process as a form of package unit in the control of the emissions from small stationary combustion sources requires consideration of a number of diverse factors which must be related and compared in a meaningful fashion. The approach taken to this evaluation involves the following steps.

(1) Development of evaluation criteria
(2) Evaluation of each stack gas cleaning process with respect to each criterion.

The conversion of the evaluation to a rating scale would be desired for the rating of the sorption processes based on the aggregate points. However, the procedure involves subjective judgements which would influence the outcome significantly. The quantitative analysis of the evaluation, therefore, was not conducted in this evaluation.

Evaluation Criteria

The characteristics of FGD sorption processes that are important in determining the packageability of such systems for small-scale nonutility combustion sources include:

(1) Size
(2) Installation
(3) Raw material availability
(4) Operation - maintenance
(5) Residual and secondary emissions
(6) By-products
(7) Capital requirement
(8) Annual cost
(9) Process availability.

The overall size of the process is to be small enough for retrofit to the existing boiler system. The sulfur dioxide sorption processes under consideration have different configurations in design; and thus, the overall size varies from one to another. This variability is expressed in terms of size and number of components and space requirement in the installation.

The installation of a sorption process should be relatively easy so that nonspecialized personnel may be able to put the system in-place within a short period of time. This variability is evaluated with respect to type of equipment and number of component.

Materials and supplies consumed by sorption processes should be readily obtainable throughout the United States. This variability is expressed in terms of the availability.

Sorption processes installed in small boilers are to be operated by non-specialized personnel and maintained as trouble-free as possible. Operational complexity is evaluated with respect to materials handling, control requirement, and number of process steps, and the maintenance problem is projected on the basis of the number of moving parts such as pumps, valves, dampers, centrifuges, and vacuum filters, plugging and scaling possibilities, and corrosion and erosion possibilities.

The sorption processes under consideration have differing potentials for minimizing air pollutant emissions and generating new pollutant emissions. This variability is expressed in terms of residual and secondary emissions which result from the application of a sorption process. In each case, cross-media emissions (i.e., air, water, and land) are considered.

By-products resulting from the sorption processes may not be significant considering the size of unit operation and the possible type of by-products. If markets are not available for by-products, the by-products would become wastes to be disposed of.

Capital requirement indicates the amount of capital required to install a sorption process. The contribution of capital cost to annual operating cost is incorporated in the annual cost.

Annual cost consists of return on rate base, federal income tax, depreciation, and net annual operating cost. This indicates total expense of sorption process resulting from capital requirement and net operating cost.

In view of the urgency of related environmental problems, the availability of a given sorption process is an important criterion in the evaluation of its applicability. The factors of type of fuel tested, development status, and operation assessment are components of the availability consideration.

Process Evaluation

The next step in the evaluation procedure was to develop an evaluation of each process with respect to each of the nine criteria. A quantitative evaluation was employed wherever possible, otherwise qualitative categories for evaluation were developed. The evaluations with respect to capital requirement and annualized cost were not included in this section but discussed extensively in the next section. The results of this evaluation are summarized in Table 37. This summary includes 18 sorption processes.

Size and Space Requirement

The important factors in this criterion include size and number of packages in the shipment and space requirement in the installation. It is difficult to obtain accurate information for the sorption processes under consideration since many of the processes have not been commercialized yet and, consequently, little experience is available in the shipment and installation of the process systems. Therefore, in this study, the possible size of packaged systems were characterized based on the size of larger components and the number of all components excluding off-site facilities such as storage tanks, grinding machines, and additional facilities necessary to process the by-product into a more marketable form, i.e., an acid plant.

TABLE 37. SORPTION PROCESS EVALUATION MATRIX

Sorption Process	Size and Space Requirement			Ease of Installation	Raw Material Availability	Operation-Maintenance						
	Size of Vessel	Number of Component	System Area			Number of Processes	Technical Expertise	Material Handling	Plugging Scaling	Erosion Corrosion	Number of Moving Parts	Operating Temperature
Limestone slurry (Peabody)	Medium	Moderate	Moderate	Moderate	Available	5	Technician	Slurry	High	Moderate	Medium	Low
Lime scrubbing (Bahco)	Medium	Moderate	Moderate	Moderate	Available	4	Technician	Slurry	High	Moderate	Medium	Low
Double Alkali (GM)	Medium	Moderate	Moderate	Moderate	Available	6	Technician	Liquid	Moderate	Low	Medium	Low
Double Alkali (FMC)	Small	Moderate	Small	Moderate	Available	5	Technician	Liquid	Minimum	Low	Medium	Low
MgO Process (central regeneration)	Medium	Moderate	Small	Easy	Available	5	Technician	Slurry	Moderate	Moderate	Medium	High
Chemiebau	Large	Low	Moderate	Moderate	Questionable	5	Engineer	Solid	Minimum	Low	Medium	High
Foster Wheeler	Large	Moderate	Moderate	Difficult	Available	4	Engineer	Solid	Minimum	Low	Medium	High
Westvaco	Large	Moderate	Moderate	Difficult	Questionable	4	Engineer	Solid	Minimum	Low	Medium	High
Sulfacid	Large	Low	Moderate	Easy	Available	3	Technician	Liquid	Minimum	Moderate	Low	Low
Chiyoda	Large	Moderate	Large	Moderate	Available	7	Technician	Liquid	Minimum	Moderate	High	Low
Ammonia (Peabody)	Medium	Moderate	Moderate	Moderate	Questionable	8	Engineer	Liquid	Minimum	Low	High	Low
Ammonia (Catalytic, Inc.)	Medium	High	Large	Difficult	Questionable	7	Engineer	Liquid	Minimum	Low	High	High
MgO Process (Integrated)	Medium	High	Large	Difficult	Available	9	Engineer	Slurry	Moderate	High	High	High
Wellman-Lord	Medium	High	Large	Difficult	Questionable	10	Engineer	Liquid	Minimum	Moderate	High	High
Shell FGD	Medium	High	Large	Difficult	Questionable	8	Engineer	Gas	Minimum	Moderate	High	High
Citrate	Medium	High	Large	Difficult	Questionable	8	Engineer	Liquid	Minimum	Moderate	High	High
Calsox	Medium	Moderate	Moderate	Moderate	Available	6	Technician	Liquid	Moderate	Low	Medium	Low
Aqueous Carbonate	Medium	High	Large	Difficult	Available	6	Engineer	Solid	Minimum	Moderate	High	High

TABLE 37. SORPTION PROCESS EVALUATION MATRIX (Continued)

Sorption Process	Emissions			Process Availability		
	SO_2 Removal Efficiency, percent	Secondary Emissions	By-Products	Type of Fuel Tested	Development Status	Operation Assessment
Limestone slurry (Peabody)	70-90	Sludge	None	Coal	Commercial	Poor-fair
Lime scrubbing (Bahco)	70-95	Sludge	None	Oil	Commercial	Fair
Double Alkali (GM)	90	Sludge	None	Coal	Commercial	Fair
Double Alkali (FMC)	90	Sludge	None	Coal	Commercial	Fair
MgO Process (central regeneration)	80-90	None	None	Coal	Commercial	Fair
Chemiebau	80-95	None	Sulfur Activated Carbon	Coal	Pilot	Poor
Foster Wheeler	85-95	None	Sulfur	Coal	Prototype	Good
Westvaco	90	None	Sulfur	Oil	Pilot	Good
Sulfacid	80-90	Sludge	None	H_2SO_4 Plant	Prototype	Good
Chiyoda	85-95	Chloride purge	Gypsum	Oil	Commercial	Good
Ammonia (Peabody)	90	None	Ammonium Sulfate	H_2SO_4 Plant	Commercial	Fair
Ammonia (Catalytic, Inc.)	90	None	Sulfur	Oil	Pilot	Fair
MgO Process (Integrated)	80-90	None	98% H_2SO_4	Coal	Commercial	Fair
Wellman-Lord	90	Sulfate purge	Sulfur	Oil	Commercial	Good
Shell FGD	75-90	Sulfate purge	Sulfur	Oil	Commercial	Good
Citrate	85-99	Sulfate purge	Sulfur	Coal	Pilot	Good
Calsox	90	Sludge	None	Coal	Pilot	Good
Aqueous Carbonate	90	None	Sulfur	Synthetic flue gas	Pilot	Fair

In assessing the space requirement a significant factor is the area devoted to waste disposal facilities such as ponds or pits. These facilities can contribute substantially to the total system area, and the area is frequently larger than the required area for the sorption process itself. In this study all processes requiring a pond for sludge or liquid disposal were modified by adding surge or neutralizing tanks with filters for the transformation into a cake.

Installation

The ease of installation depends on type and number of equipment components and process flow in the overall operation. This criterion was qualitatively evaluated with respect to three categcries defined as follows:

Category 1 - Relatively easy to install, simple and uncomplicated process, and moderate field fabrication required.

Category 2 - Moderately difficult to install, more involved process, and more field fabrication than in Category 1.

Category 3 - Difficult to install, complex process with more process steps and extensive piping, and considerable field fabrication required.

Raw Material Availability

The availability of raw material required was evaluated on the basis of two categories defined as follows:

Category 1 - Materials readily available and in surplus generally throughout the United States, for example, lime, limestone, magnesium oxide, lignite, and soda ash.

Category 2 - Materials either in short supply or available only to specific areas, for example, hydrogen, ammonia, and methane.

Operation - Maintenance

The operational and maintenance difficulties were assessed in terms of number of process steps, technical expertise, characteristics of materials handling, possibilities of plugging, scaling, erosion, and corrosion, number of moving parts, and operating temperatures.

The number of process steps is based on the number of operation or process steps that require monitoring. This, in general, includes adsorption, regeneration, evaporation, stripping, thickening, centrifuging, heating, drying, and other related processes.

The degree of technical expertise required is based on the process complexity, control sensitivity, and operating conditions. It was expressed in terms of technical knowledge equivalent to either technician or engineer.

The material handling was evaluated in terms of gases, liquids, solids, and slurry. The handling of solids or slurry is more difficult than that of liquids or gases in transportation and processing.

Some sorption processes involve solid or slurry streams that are more susceptible to scaling and/or plugging than others. Scaling and plussing can precipitate equipment failure and result in operational disruptions. The potential was evaluated with respect to three categories as follows:

Category 1 - Minimal possibility
Category 2 - Moderate possibility
Category 3 - High possibility.

In SO_2 sorption processes, corrosion is caused primarily by the presence of dilute sulfuric acid and/or chlorine ions. Erosion is caused by the abrasive nature of liquids and solids. Both corrosion and erosion were evaluated with respect to three categories:

Category 1 - Minimal possibility
Category 2 - Moderate possibility
Category 3 - High possibility.

The number of moving parts was the summation of all of the major pieces of equipment containing moving parts. This included converyors, rotary drum filters, pumps, blowers, mixers, etc. This factor was categorized in terms of low, moderate, and high.

Operating temperatures influence the reliability of process operation to some extent. A high operating temperature is more conclusive to failure than a low one.

Residual and Secondary Emissions

The residual emission of sulfur dioxide was evaluated on the basis of the removal efficiency. The control of fly ash emission from the existing control system was not taken into consideration in this study except for the dry limestone injection process. The secondary emissions resulting from the sulfur dioxide control process were expressed in terms of pollutant and the quantity per pound of SO_2 removed.

By-Products

The by-products were evaluated with respect to marketability and ease of handling. The following three categories were employed:

Category 1 - Moderate marketability and easy to handle: elemental sulfur activated carbon

Category 2 - Moderate marketability and relatively difficult to handle: sulfuric acid

Category 3 - Poor marketability and easy to handle: gypsum, ammonium sulfate

Process Availability

The process availability was evaluated on the basis of type of fuel tested, development status, and operation assessment. The development status described the state of development of each system as a nonpackage unit. It was classified into four categories - bench, pilot, prototype, and commercial. The operation assessment indicating the degree of successful operation was classified into three categories - good, fair, and poor.

COST OF SORPTION PROCESSES

The capital requirement and annualized cost of the various sorption processes under consideration were estimated for the boiler subgroups of environmental concern described in Part II of this report. Since the concept of the package unit may not be feasible for the large size class (i.e., the boiler size of 250,000 lb/hr) probably due to the size limitation, the MUC source class was excluded in the cost estimation.

The capital requirement employed in this study included costs for equipment, material, installation, engineering and design, and startup (battery limit cost). Although the capital requirement for a package unit was preferred in this study, it was very difficult to obtain such information because of lack of data. The base year for the estimation was mid-1973 and the Utility Financing Method listed in Appendix A was employed to estimate the related costs.

The annualized cost included fixed capital charges, labor, utilities, raw material, and by-product credit. The cost was estimated based on the same format used in Part II of this report. The following assumptions were made:

(1) The existing coal-fired boiler system has been equipped with a flyash collecting system. The Shell FGD System would need a more efficient ESP system.

(2) The sludge generated from the throwaway processes would be filtered and disposed of in landfills.

(3) Flue gases from FGD processes would be reheated, if necessary, using an indirect steam reheat system.

(4) The retrofit factor was assumed to be 1.2 for the double alkali systems, 1.4 for Shell FGD systems, and 1.3 for other sorption processes.

(5) The costs for oil-fired boilers were estimated from those for coal-fired boilers with adjustments made with respect to flue gas flow rate and sulfur input.

The summary of the estimations is shown in Tables 38 and 39 for coal- and oil-fired boilers, respectively. Among the various sorption processes, the MgO process with regeneration performed at a central facility appears most attractive economically. Both the capital requirement and annualized cost are relatively low compared with those for other sorption processes. Throwaway processes in general are low in capital requirement and annualized cost than regenerable processes with on-site regeneration facilities.

TABLE 38. CAPITAL REQUIREMENT AND ANNUALIZED COST OF SORPTION PROCESSES FOR COAL-FIRED BOILERS

C: Commercial Boiler I: Industrial Boiler

Sorption Process	Capital Requirement, $\$10^3$		Annualized Cost			
			Steam Output, $\$/10^6$ Btu		Sulfur Removal, $/lb S	
	20 (C)	20 (I)	20 (C)	20 (I)	20 (C)	20 (I)
Limestone slurry (Peabody)	616	624	2.04	1.71	0.87	0.73
Lime scrubbing (Bahco)	780	780	2.50	2.04	1.00	0.82
Double alkali (GM)	754	761	2.54	2.10	0.96	0.79
Double alkali (FMC)	697	702	2.27	1.86	0.86	0.70
MgO (central regeneration)	482	490	1.69	1.41	0.64	0.53
Chemiebau	971	979	3.23	2.64	1.38	1.12
Foster Wheeler	810	818	2.84	2.37	1.07	0.89
Westvaco	1,136	1,152	4.00	3.34	1.51	1.26
Sulfacid	1,002	1,013	3.17	2.59	1.20	0.98
Chiyoda	879	887	2.95	2.49	1.11	0.93
Ammonia scrubbing (Peabody)	759	770	2.49	2.02	0.94	0.76
Ammonia scrubbing (Catalytic, Inc.)	1,117	1,125	3.91	3.23	1.48	1.22
MgO (integrated)	858	871	3.32	2.81	1.25	1.06
Wellman-Lord	757	765	2.76	2.29	1.04	0.87
Shell FGD	924	929	3.18	2.52	1.20	0.93
Citrate	806	816	2.60	2.10	0.98	0.79
Calsox	604	609	2.27	1.84	0.86	0.70
Aqueous carbonate	1,300	1,300	4.40	3.61	1.66	1.36

TABLE 39. CAPITAL REQUIREMENT AND ANNUALIZED CONTROL COST OF SORPTION PROCESSES FOR OIL-FIRED BOILERS

C: Commercial Boiler I: Industrial Boiler

Sorption Process	Capital Requirement, $\$10^3$				Annualized Cost							
					Steam Output, $\$/10^6$ Btu				Sulfur Removal, \$/lb S			
	2 (C)	2 (I)	20 (C)	20 (I)	2 (C)	2 (I)	20 (C)	20 (I)	2 (C)	2 (I)	20 (C)	20 (I)
Limestone slurry (Peabody)	139	139	498	503	5.32	4.91	2.26	1.89	4.40	4.06	1.87	1.56
Lime scrubbing (Bahco)	173	173	637	637	7.53	6.89	2.85	2.30	5.84	5.34	2.21	1.78
Double alkali (GM)	162	162	580	584	6.56	6.09	2.75	2.27	4.83	4.48	2.03	1.67
Double alkali (FMC)	154	154	538	540	6.85	6.34	2.47	2.01	5.04	4.47	1.82	1.48
MgO (central regeneration)	111	111	393	398	4.78	4.46	1.90	1.60	3.52	3.29	1.40	1.18
Chemiebau	--[a]	--	676	681	--	--	3.24	2.68	--	--	2.68	2.21
Foster Wheeler	--	--	607	612	--	--	3.06	2.52	--	--	2.25	1.86
Westvaco	--	--	751	759	--	--	3.89	3.28	--	--	2.87	2.42
Sulfacid	222	222	803	806	7.86	7.20	3.61	2.92	5.79	5.30	2.66	2.15
Chiyoda	190	190	677	683	7.86	7.31	3.24	2.68	5.79	5.38	2.39	1.97
Ammonia scrubbing (Peabody)	--	--	527	530	--	--	2.53	2.14	--	--	1.86	1.58
Ammonia scrubbing (Catalytic, Inc.)	--	--	735	740	--	--	3.75	3.16	--	--	2.76	2.32
MgO (integrated)	--	--	607	612	--	--	3.30	2.79	--	--	2.43	2.05
Wellman-Lord	--	--	602	607	--	--	3.08	2.58	--	--	2.27	1.90
Shell FGD	--	--	876	879	--	--	4.34	3.46	--	--	3.20	2.54
Citrate	--	--	620	625	--	--	3.14	2.60	--	--	2.31	1.91
Calsox	122	123	591	594	5.44	5.03	3.08	2.49	4.01	3.70	2.27	1.83
Aqueous carbonate	--	--	791	791	--	--	4.00	3.30	--	--	2.94	2.43

(a) The blank indicates that the process is too complicated to be applied to the size class.

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(101) Pohlenz, J. B., "The Shell Flue Gas Desulfurization Process," a paper presented at the Flue Gas Desulfurization Symposium, Atlanta, Georgia, November 4-7, 1974.

(102) Private communication between J. B. Pohlenz of Union Oil Products and P. Choi of Battelle-Columbus, January 6, 1975.

(103) Dantzenberg, F. M., Naber, J. E., and Van Ginneken, A. J. J., "The Shell Flue Gas Desulfurization Process," a paper presented at the 68th National Meeting of AIChE, Houston, Texas, February 28 - March 4, 1971.

(104) Personal communication with R. E. Barnard, and Richard Teague, Monsanto Enviro-Chem Systems, Inc., St. Louis, Missouri, September 24, 1974.

(105) Botts, W. V., and Gehri, D. C., "Regenerative Aqueous Carbonate Process (ACP) for Utility and Industrial SO_2 Removal Applications," presented at 167th American Chemical Society National Meeting, Los Angeles, California, April 4, 1974.

APPENDIX A

ACCOUNTING METHOD

APPENDIX A

ACCOUNTING METHOD

An accounting method was derived from the Utility Financing Method as modified by the Panhandle Eastern Pipeline Company. The accounting method was then employed uniformly for all cost estimations. A description of the method follows.

Total Plant Investment and Total Capital Requirement

Total Bare Cost

The total bare cost includes major equipment costs, direct construction labor costs, undistributed costs such as costs for construction facilities and services, and other plant costs such as for utilities and off-site facilities. This cost is used as a basis for other cost estimations.

Engineering and Design Cost

This cost generally is assumed at 5 percent of the total bare cost. In some available data, this cost was included in the total bare cost.

Contractor's Overhead and Profits

This cost is assumed at 10 percent of the total bare cost. In some available data, this cost was included in the total bare cost.

Subtotal Plant Investment

This is the summation of the total bare cost, engineering and design cost, and contractor's overhead and profits.

Project Contingency

This represents the degree of uncertainty in the overall construction cost estimate. This is assumed at 15 percent of the subtotal plant investment.

Total Plant Investment (TPI)

This is the summation of the subtotal plant investment and project contingency.

Interest During Construction (IDC)

This is obtained from the following equation.

IDC = (interest rate)(TPI)(average construction period in year).

Startup Cost

This cost is assumed at 20 percent of the gross annual operating cost.

Working Capital (WKC)

This cost is also assumed at 20 percent of the gross annual operating cost.

Total Capital Requirement (TCR)

The total capital requirement is given as the summation of the total plant investment, interest during construction, startup cost, and working capital.

Net Annual Operating Cost

Direct Material and Utilities

This cost includes costs for raw materials consumed in the process and utilities such as power, fuel, process steam, process and cooling water, and compressed air.

Maintenance and Operating Supplies

This cost includes costs of supplies for operating and maintenance. The cost is assumed to be the summation of 30 percent of direct operating labor and 1.5 percent of total plant investment, if not specified.

Direct Operating Labor (DOL)

This cost is obtained using the following equation.

DOL = (man-hour required/hr)($5/man-hour)(8,304 hrs/yr).

Maintenance Labor

The annual maintenance labor cost is given as 1.5 percent of the total plant investment, if not specified.

Supervision

The cost for supervision in general is assumed at 15 percent of the summation of direct operating labor and maintenance labor costs.

Administration and General Overhead

This cost is assumed at 60 percent of the total labor cost including supervision.

Local Taxes and Insurance

The annual local taxes and insurance are estimated at 2.7 percent of the total plant investment.

Gross Operating Cost

This is the summation of all annual operating costs listed above.

By-Product Credit

This credit comes from the sale of by-products. The by-product credit is subtracted from the gross operating cost to obtain the net operating cost.

Net Annual Operating Cost (AOC)

The net annual operating cost is obtained by subtracting the total by-product credit from the gross annual operaring cost. The escalation of net annual operating cost was not considered in this study.

Annualized Cost

The control cost represents the average cost during the life of the plant. The basis for the calculation is:

- 20-year plant life
- 5 percent per year straight line depreciation on total capital requirement excluding working capital

- 75 percent/25 percent debt/equity ratio
- 9 percent per year interest on debt
- 15 percent per year return on equity after tax
- 48 percent Federal income tax rate.

For retrofit processes such as fluidized combustion and FGD processes, the life of the system was assumed at 10 years and accordingly, 10 percent per year straight line depreciation was used. The following procedures are used to calculate the average annual revenue requirement and product (or control) cost (for the case of 5 percent straight line depreciation).

Average Annual Depreciation (D) = 0.05 (TCR-WKC).

Average Return on Rate Base (RRB) = 0.0525 (TCR + WKC).

Average Federal Income Tax (FIT) = 0.01731 (TCR + WKC).

Average Annual Revenue Requirement (ARR) = RRB + FIT + D + Net AOC.

		Unit Cost	
Coal, \$/ton		10	
Limestone, \$/ton (pulverized limestone)		7	(10)
Lime, \$/ton		25	
NaOH, \$/ton		150	
$NaCO_3$, \$/ton		50	
MgO, \$/ton		140	
Ammonia, \$/ton		150	
Fuel oil, \$/bbl (high S)		3	
No. 2 fuel oil, \$/bbl		4.5	
Carbon dioxide, \$/ton		60	
H_2SO_4 (98 percent), \$/ton	By-product credit	20	
H_2SO_4 (80 percent), \$/ton		10	
Elemental sulfur, \$/ton		10	
Labor, engineer, \$/hr		7.5	
technician, \$/hr		5	
Electricity, \$/kwhr		0.01	
Process water, $\$/10^3$ gal		0.5	
Cooling water, $\$/10^3$ gal		0.1	
Steam, $\$/60^6$ Btu		0.5	
Ferric sulfate, \$/ton		50	
Coke, \$/ton		40	
Oxygen, \$/ton		10	
Sand, \$/ton		5	
Activated carbon, \$/ton		800 (for Westvaco) 300 (for FW) 60 (for by-product sale)	

APPENDIX B

DESCRIPTION OF FLUE GAS DESULFURIZATION PROCESSES

APPENDIX B

DESCRIPTION OF FLUE GAS DESULFURIZATION PROCESSES

Limestone Slurry (Peabody Engineering)

Developer/Manufacturer

Peabody Engineering Systems, Stamford, Connecticut, has developed, and is currently marketing a limestone based SO_2 removal system for industrial boilers.

Process Description

The Peabody process utilizes a lime, limestone, or soda ash slurry in a spray tower absorber to remove SO_2 and particulates from flue gas. For the limestone system, the spent slurry is thickened and filtered and the filter cake, containing predominantly $CaSO_4$, is trucked away. The overflow is returned to the absorber to complete the loop.

The flue gas, after passing through a multiple cyclone separator or ESP, is passed through an ID fan and into the quench system where it is cooled by a contacting spray of limestone slurry (see Figure B-1). It flows through the absorber, a three or five bank spray tower, where the SO_2 reacts with the limestone slurry to form $CaSO_3$ and $CaSO_4$. In the spray tower, 70 to 90 percent of the $CaSO_3$ is oxidized to $CaSO_4$. After passing through an impingement tray and a mist eliminator the gas exits the scrubber, and flows through a reheater, and is vented to the stack. The system operates with a liquid-to-gas ratio of 75-100 gal/1000 ft^3. Pressure drop through the tower is about 4 inches of water.

$$\begin{aligned}
SO_2 + H_2O &= H_2SO_3 \\
H_2SO_3 + CaCO_3 &= CaSO_3 + H_2O + CO_2 \\
CaSO_3 + \tfrac{1}{2}\,O_2 &= CaSO_4 \\
CaSO_3 + \tfrac{1}{2}\,H_2O &= CaSO_3 \cdot \tfrac{1}{2}\,H_2O \\
CaSO_4 + 2H_2O &= CaSO_4 \cdot 2H_2O
\end{aligned}$$

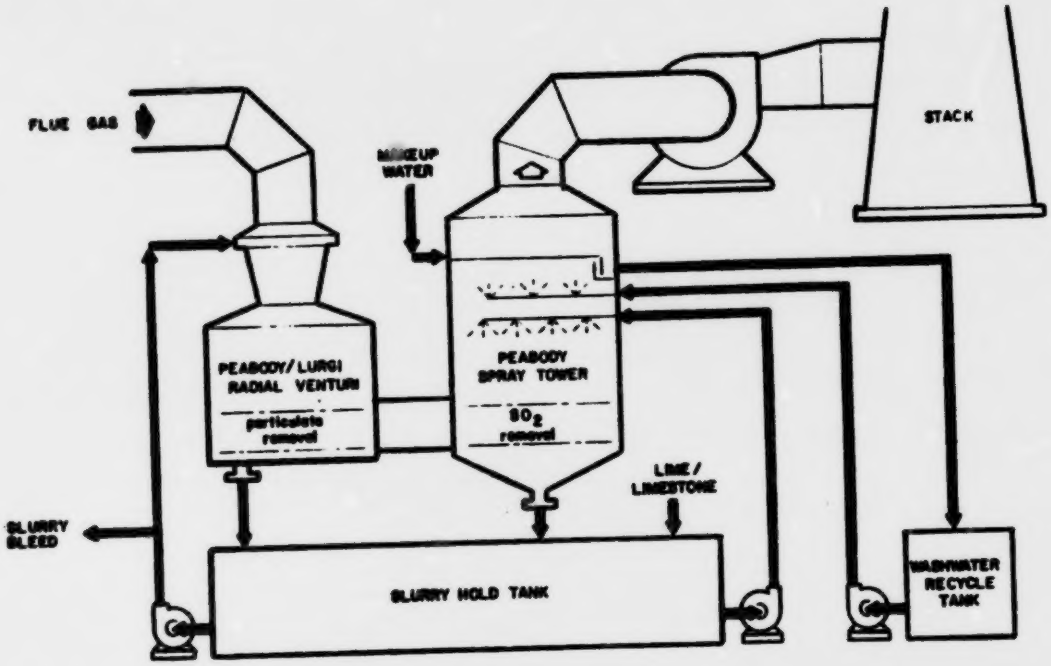

FLUE GAS
MAKEUP WATER
STACK
PEABODY/LURGI RADIAL VENTURI
particulate removal
PEABODY SPRAY TOWER
SO_2 removal
LIME/ LIMESTONE
SLURRY BLEED
SLURRY HOLD TANK
WASHWATER RECYCLE TANK

FIGURE B-1. PEABODY LIMESTONE SCRUBBING PROCESS

The spent slurry drains from the bottom of the tower into the recycle tank. Limestone is added to the recycle tank to control the pH. Slurry overflows from the tank to a sump tank from where it is pumped to a thickener. The $CaSO_4$ crystals are preferred over $CaSO_3$ cyrstals because of their better settling and dewatering characteristics. This is why the high oxidation rate in the spray tower is desired. The thickener underflow is pumped to a vacuum filter where a 70 percent solids cake is formed. The cake is conveyed to a truck and transported to a landfill area. Thickener overflow and the filtrate are returned to the recycle tank to maintain closed loop operation.

Removal Efficiencies

The quench-tower system is capable of a 70 to 90 percent SO_2 and 95 percent fly ash removal efficiency. For higher fly ash removal efficiencies, 99 percent, a Peabody Radial Flow Venturi may be substituted for the quench system.

Wastes

The only waste stream is the filter cake containing predominantly $CaSO_4$. About 4 lb of filter cake (70 percent solids) is generated per pound of SO_2 removed.

By-Products

None.

Materials of Construction

All slurry pumps and flow lines are rubber lined. The reaction tank and scrubber are lined with fiberglass reinforced polyester. The absorber interface tray is constructed of 316 L stainless steel.

Raw Material and Heat Requirements

The only raw material required is pulverized limestone. Heat is required to reheat the flue gas.

Advantages and Disadvantages

Advantages:
(1) Simple operation
(2) Low pressure drop
(3) Low cost absorbent (limestone)
(4) Good turndown ratio
(5) Comparatively low operating cost.

Disadvantages:
(1) Waste sludge generation
(2) Stack gas reheat may be required
(3) High liquid to gas ratio.

Development Status

In 1973 Peabody designed and built a 1-MW pilot plant SO_2 scrubbing facility at Detroit Edison's River Rouge Station, Detroit. It has reportedly been operating satisfactorily since startup in February, 1974. A full-scale facility is presently undergoing start-up at Detroit Edison's St. Clair Unit No. 6, a 175-MW coal-fired boiler.

Capital and Operating Costs

The 1974 capital cost for a Peabody limestone SO_2 scrubbing system was estimated at $0.8 million and $1.4 million for coal-fired 50,000 and 150,000 lb steam/hr boiler systems, respectively. The cost includes the engineering and design cost of the system including the reheater. Table B - 1 shows the labor, material, and utility requirements of the Peabody system installed on a 150,000 lb steam/hr boiler system.

TABLE B-1. LABOR, MATERIALS, AND UTILITY REQUIREMENTS FOR A PEABODY LIMESTONE SCRUBBING PROCESS(55)

Basis: 150,000 lb steam/hr coal boiler
3 percent sulfur in coal
75 percent load factor
6,500 hours/yr operation
80 percent removal efficiency
92,000 acfm of flue gas at 450°F

Item	Quantity
Utility	
Power	330 kW
Steam	2,800 lb/hr
Water	50 gpm
Material	
Limestone	0.9 tons/hr
Labor	
Direct operation	1 man/shift
Maintenance-labor	1/3 man shift
Material	2 percent of capital cost/yr

Lime Scrubbing (Bahco)

Developer/Manufacturer

The Bahco SO_2 removal process was developed by A. B. Bahco Ventilation, Enkoping, Sweden. The process is being marketed in the United States by Research-Cottrell, Inc., Bound Brook, New Jersey.

Process Description

The Bahco process used a lime slurry in a two-stage venturi scrubber to remove particulate matter and SO_2 from flue gas.(56, 57) As can be seen in Figure B - 2 , flue gas is introduced into the bottom stage of the scrubber where it contacts hydrated lime slurry and passes into the first venturi. Sulfur dioxide reacts with the lime slurry to form calcium sulfite and calcium sulfate.

$$Ca(OH)_2 + SO_2 + H_2O \rightarrow CaSO_3 \cdot 2H_2O$$

$$CaSO_3 \cdot 2H_2O + \frac{1}{2} O_2 \rightarrow CaSO_4 \cdot 2H_2O$$

The atomized droplets are separated from the gas at the top of the venturi by a centrifugal force drop collector. The gas is directed to the second venturi and the collected liquid is returned to the first stage contact zone. A concentration regulator continuously withdraws a fraction of return stream and feeds it to a thickener or a drum filter. After contacting the second stage drop collector, the gas is sent through a reheater where it is heated to 175°F and expelled to the stack. The collected liquid is also returned to the first stage contact zone where any overflow is pumped to the dissolving tank and ultimately returned to the second stage impingement zone. The level in the first stage is regulated

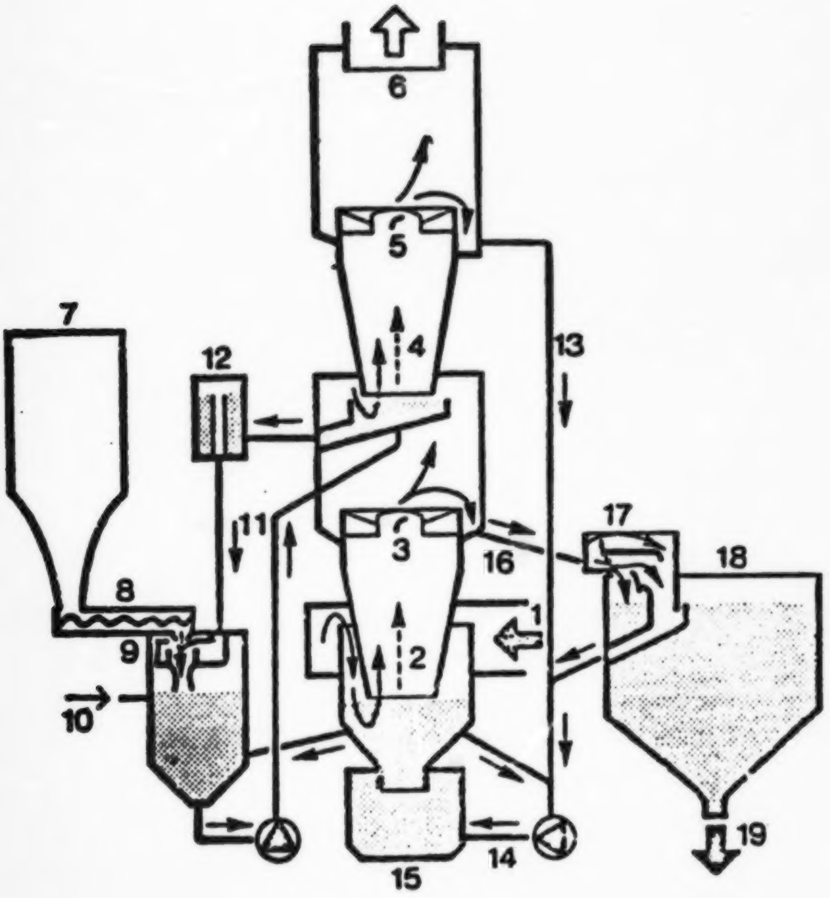


FIGURE B - 2 . BAHCO SO_2 SCRUBBER(56)

1. Flue Gas Inlet
2. First Scrubber Stage
3. Guide Vane
4. Second Scrubber Stage
5. Guide Vanes
6. Clean Gas Outlet
7. Storage Bin
8. Screw Feeder
9. Dissolving Tank
10. Water Supply
11. Tubes
12. Level Tank
13. Tubes
14. Tubes
15. Sludge Mill
16. Tubes
17. Concentration Regulator
18. Settling Tank
19. Sludge Outlet

by controlling the level in the dissolving tank. The sludge stream leaving the thickener is concentrated in a drum filter.

Removal Efficiency

The sulfur dioxide removal efficiency ranges from 70 percent to 99 percent depending on sulfur dioxide concentration in flue gas.

Raw Materials

Materials required in Bahco process include lime and process water.

By-Product

No by-product is obtained from the Bahco process.

Wastes

Waste emissions from the Bahco process include sulfur dioxide residual emission and sludge resulting from the process. About 5.1 lbs of calcium sulfate sludge (50 percent solids) are produced per pound of SO_2 removed.

Advantages

(1) Simple process
(2) Simultaneous removal of fly ash and sulfur dioxide
(3) High SO_2 removal efficiency
(4) High reliability.

Disadvantages

(1) High lime cost
(2) Waste sludge generation
(3) High power requirement
(4) Reheating requirement for flue gas.

Development Status

In 1964, A. B. Bahco Ventilation of Sweden initiated investigations of sulfur dioxide control using alkaline base scrubbing reagents. In 1966, they installed a 1,400 scfm pilot unit on the boiler of their central heating plant. In 1969, their first commercial unit was installed on an oil-fired boiler producing 75,000 lbs/hour of steam. In 1970, Bahco licensed their process technology rights in Japan to Marubeni with Tsukishima Kikai as a sublicensee. In August, 1971, Research-Cottrell acquired the rights to license the Bahco system in the United States and Canada. Currently, 19 commercial units have been installed in Japan and Sweden, and one unit will be installed on a coal-fired boiler in the United States in 1975. Table B-2 summarizes the location and service of these installations.

Capital and Operating Costs

The installed cost for a Bahco system using carbon steel and treating the flue gas characterized in Table B-3, was estimated at $0.8 million in 1971.(56) This did not include any unique installation costs such as interconnecting duct work, utility connections, remote instrumentation. Total installed system costs are often significantly higher than installed costs. For example, for a Bahco system to be installed on a 18-MW coal-fired stoker boiler at Rickenbacker Air Force Base, Columbus, Ohio, the total installed cost was quoted as

TABLE B-2 . INSTALLATIONS OF THE BAHCO SO_2 REMOVAL SYSTEM(57)

Company	Location	No. of Units	Unit Capacity, scfm at 32 F	Service	Scrubbing Reagent	Inlet SO_2 Conc., ppm	SO_2 Removal Efficiency, percent
Sodersjukhuset	Stockholm, Sweden	3	17,700	Oil-fired boiler	$Ca(OH)_2$	800-1500	97-99
Daishowa Seishi Co.	Suzukawa, Japan	1	14,700	Oil-fired boiler	NaOH	900-1200	97-99
Daishowa Seishi Co.	Yoshinaga, Japan	5	44,200	Oil-fired boiler	NaOH	900-1000	97.5
Osaka City	Osaka, Japan	1	10,000	Secondary sludge incinerator	NaOH		
Hiroshima City	Hiroshima, Japan	1	10,000	Secondary sludge incinerator	NaOH		
Yahagi Iron Works	Nagoya, Japan	1	48,300	Sintering plant	$Ca(OH)_2$ waste carbide sludge	2500-4000	90-95
Taio Paper Co.	Iyomishima, Japan	1	83,000	Oil-fired boiler	NaOH	1000-1500	98
Central Glass Co.	Sakai, Japan	1 3	31,300	Glass furnace	NaOH	1200 (25% SO_2)	98
Stora Kopparberg	Gryeksbo, Sweden	1	17,700	Black liquor boiler	CaO and $CaCO_3$ dust	1000-6000	70
Kanegafuchi Chemical	Takasago City, Japan	2	79,500	Oil-fired boiler	NaOH		
Daishowa Seishi Co.	Yoshinaga, Japan	2	66,400	SCA-Billerud recovery boiler	NaOH		
Rickenbacker Air Force Base(a)	Columbus, Ohio, United States	1	45,000	Coal-fired boiler	$Ca(OH)_2$		

(a) Start-up late 1975.

TABLE B - 3. FLUE GAS CHARACTERISTICS FOR BAHCO PROCESS(56)

Item	Value
Flow rate	137,000 acfm
Temperature	350° F
SO_2 concentration	2,000 ppm
Fly ash loading	2.4 gr/scf

about $2.3 million including costs for dust collector, sludge pond, etc., in 1974. The system will be based on stainless steel and the construction will begin in early 1975. The exponential scale-up factor for equipment cost was quoted as 0.6-0.7.

Operating costs consist of labor, material, and utilities. The requirements for treating the flue gas characterized in Table B-3 are shown in Table B-4.

TABLE B-4. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR BAHCO PROCESS(56)

Item	Quantity
Utilities	
Power	600 kW
Makeup water	40 gpm
Material	
Lime	0.94 tons/hr
Labor	
Direct operation	0.5 man/shift
Maintenance*	3 percent of total plant investment/yr

* This includes labor and material.

Double Alkali Process (FMC)

Developer/Manufacturer

The FMC/Link-Belt Double Alkali SO_2 Absorption System was developed by FMC Corporation.

Process Description

In an effort to overcome the scaling difficulties with direct calcium slurry scrubbing, FMC employs a "concentrated" sulfite/bisulfite buffer solution to remove particulate matter and SO_2 from flue gas. The spent solution adds lime or limestone outside of the scrubber system.(58)

A simplified flow sheet of the FMC/Link-Belt Process is shown in Figure B - 3 .(59) The flue gas enters the scrubber at about 300°F after passing through a cyclone dust collector and a forced-draft fan. The scrubber is a dual throat variable-flow venturi where both fly ash and SO_2 are removed by contacting with a 20 weight % solution of Na_2SO_3, $NaHSO_3$, and Na_2SO_4. Sulfur dioxide reacts with the sodium sulfite to form sodium bisulfite; some sodium sulfite oxidizes to sodium sulfate.

$$SO_3^{=} + SO_2 + H_2O \rightarrow 2HSO_3^{-}$$

$$SO_3^{=} + 1/2\ O_2 \rightarrow SO_4^{=}$$

The scrubber normally operates with a liquid-to-gas ratio of 10 $gal/10^3\ ft^3$.

In the cyclone, the water droplets separate from the gas and descend to the bottom. The flue gas exits through the top of the scrubber, is reheated to 200°F and is sent to the stack. The spent slurry is pumped from the bottom of the scrubber to the recirculation tank. A regeneration stream is withdrawn from the recycle stream at a rate equal to the rate of SO_2 collected in the scrubber. It is transferred to the lime reactor where lime is added to form insoluble $CaSO_3$ and regenerate the Na_2SO_3.

$$SO_3^{=} + Ca(OH)_2 \rightarrow CaSO_3\downarrow + 2OH^{-}$$

$$2OH^{-} + 2HSO_3 \rightarrow 2SO_3^{=} + 2H_2O$$

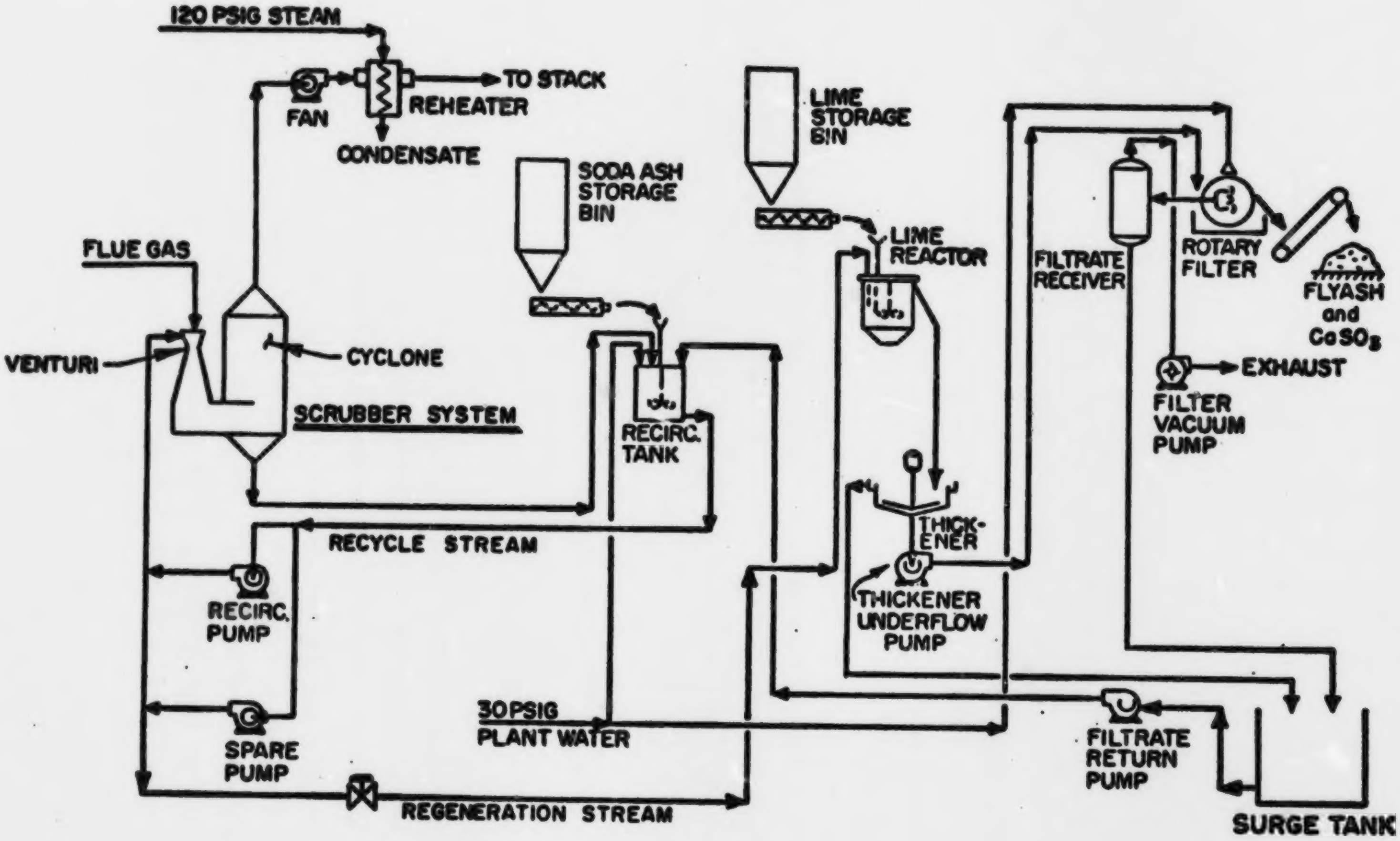

120 PSIG STEAM
TO STACK
FAN
REHEATER
CONDENSATE
SODA ASH STORAGE BIN
LIME STORAGE BIN
LIME REACTOR
FLUE GAS
VENTURI
CYCLONE
SCRUBBER SYSTEM
RECIRC. TANK
ROTARY FILTER
FILTRATE RECEIVER
FLYASH and $CaSO_3$
EXHAUST
FILTER VACUUM PUMP
THICK-ENER
THICKENER UNDERFLOW PUMP
RECYCLE STREAM
RECIRC. PUMP
SPARE PUMP
30 PSIG PLANT WATER
REGENERATION STREAM
FILTRATE RETURN PUMP
SURGE TANK

FIGURE B-3. FMC/LINK-BELT ALKALINE ABSORPTION PROCESS FOR SULFUR DIOXIDE CONTROL (59)

Shipping and Construction

For small units, i.e., less than 20 MW, the venturi scrubber and vacuum filter may be prefabricated in the shop. The large size of the storage tanks and thickener, however, requires that they be shipped to the construction site in sections and field fabricated.

By-Product

No marketable by-product.

Raw Materials and Heat Requirements

Lime and soda ash are the required raw materials. Steam is necessary to reheat the flue gas from 120 to 200°F.

Advantages

(1) High SO_2 and particle removal efficiency with a liquid-to-gas ratio

(2) No severe plugging or scaling problems

(3) High reliability (minimal sulfite oxidation and greater pH flexibility in scrubber).

Disadvantages

(1) Waste sludge generation

(2) Flue gas reheat required

(3) High lime cost

(4) High power requirement.

Complete lime reaction and formation of calcium sulfite is insured by maintaining a high sulfite concentration. The mixture of calcium sulfite, sodium sulfite, sodium sulfate and fly ash is transferred to the thickener. The thickener overflow, containing soluble sodium sulfite and sodium sulfate is returned to the recirculation tank. The thickener underflow, containing 25 to 30 percent solids, is transferred to a rotary vacuum filter. The resultant filter cake contains about 55 percent solids, composed primarily of $CaSO_3$ and fly ash with about 4 to 5 percent Na_2SO_3 and Na_2SO_4. Most of the sodium salts are recovered from the cake by washing and are returned with the filtrate to the recirculation tank. The sodium losses are made up by the addition of soda ash to the recirculation tank.

Removal Efficiencies

The fly ash and SO_2 removal efficiencies are 99 and 90 percent, respectively.

Wastes

The only waste emission is the $CaSO_3$ and fly ash filter cake. About 5.85 lb of solid waste (55 percent solids) is generated per pound of SO_2 removed.

Materials of Construction

The scrubber is constructed of 316L stainless steel. To prevent fly ash abrasion, slurry lines and pumps are rubber lined. Direct steam tube reheat requires Hastelloy G heating tubes. Indirect reheating of outside air and blending with the flue gas may be accomplished with carbon steel tubes.

Development Status

The chemistry of this process has been tested at FMC's 20,000 acfm barium sulfate reduction kiln in Modesto, California. A packed bed absorber is used instead of the venturi to remove SO_2 with inlet concentrations of up to 8,000 ppm. FMC reports that since startup in December, 1971, the system has operated troublefree for over 22,000 hours.(59)

In 1971, a semi-trailer was fitted with a 3,500 acfm pilot plant and several runs have been made on different coal-fired boilers. It is still operating and is a demonstration device for marketing purposes.

Construction has begun on the only large industrial unit at Caterpillar Tractor's 45-MW boiler at Mossville, Illinois. Startup is expected soon.

Capital and Operating Costs

The capital cost for a coal-fired boiler (capacity equivalent to 45 MW) was estimated at $3.057 million in 1973.(59) The labor, material, and utility requirements of the system are shown in Table B-5.

TABLE B-5. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR FMC DOUBLE ALKALI PROCESS (Capacity, 45 MW)

Item	Quantity
Utility	
Power	1 MW
Steam	17,600 lb/hr
Water	70 gpm
Material	
Lime (92 percent)	1.64 tons/hr
Soda ash	0.25 tons/hr
Labor	
Direct operation	1/2 man shift
Maintenance	2 percent of capital cost

* This includes labor and material.

MgO Process

Developer/Manufacturer

Chemical Construction Corporation, New York, New York, developed the regenerable magnesium oxide-sulfur dioxide removal process.

Process Description

The flue gas at 300°F is passed through an electrostatic precipitator and is introduced into the top stage of a two-stage venturi scrubber/absorber where the fly ash and particulate matter are removed by contacting a water spray (see Figure B-4). The gas passes into the lower stage where it contacts a 12 percent solid slurry of MgO, $MgSO_3$, and $MgSO_4$. The SO_2 in the flue gas reacts with the MgO to form $MgSO_3$; some oxidation to $MgSO_4$ occurs, normally 15-20 percent.

Absorption Main Reactions:

$$MgO + SO_2 + 3H_2O \rightarrow MgSO_3 \cdot 3H_2O$$

$$MgO + SO_2 + 6H_2O \rightarrow MgSO_3 \cdot 6H_2O$$

Side Reactions:

$$MgSO_3 + SO_2 + H_2O \rightarrow Mg(HSO_3)_2$$

$$Mg(HSO_3)_2 + MgO \rightarrow 2MgSO_3 + H_2O$$

$$MgO + SO_3 + 7H_2O \rightarrow MgSO_4 \cdot 7H_2O$$

$$MgSO_3 + \frac{1}{2} O_2 + 7H_2O \rightarrow MgSO_4 \cdot 7H_2O$$

The normal liquid to gas ratio is about 33 $gal/10^3 ft^3$. At design operating conditions, the total pressure drop through the scrubber is about 6 in water.

The purified gas exits the scrubber at 125°F, is reheated, and vented to the stack.

The water-fly ash slurry drains from the venturi section of the scrubber into a split stream; one stream is recycled to the scrubber,

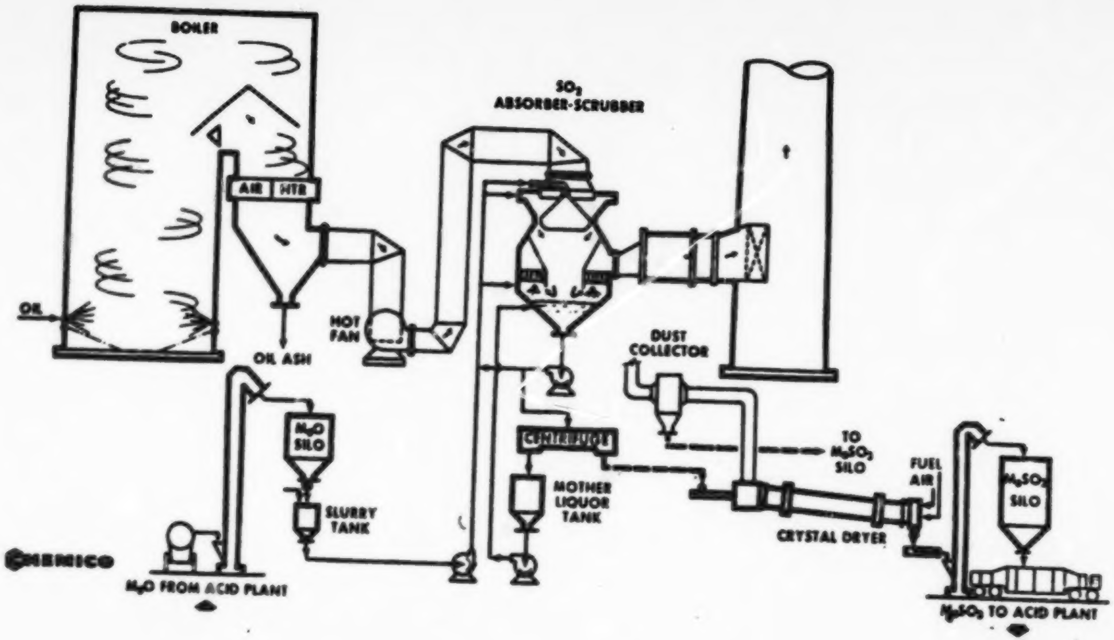

BOILER
SO_2 ABSORBER-SCRUBBER
AIR HTR
OIL
OIL ASH
HOT FAN
DUST COLLECTOR
MgO SILO
SLURRY TANK
CENTRIFUGE
MOTHER LIQUOR TANK
TO $MgSO_3$ SILO
FUEL AIR
$MgSO_3$ SILO
CRYSTAL DRYER
CHEMICO
MgO FROM ACID PLANT
$MgSO_3$ TO ACID PLANT

FIGURE B-4. MgO SCRUBBING PROCESS FLOW SHEET (60) the other flows to a thickener. The thickener bottom, containing principally fly ash with some insoluble impurities, is pumped to a settling pond; the thickener overflow is returned to the top of the scrubber.

The magnesium salt slurry drains into the sump from where a slipstream is withdrawn and pumped to a centrifuge. In the centrifuge, the solids are separated and the mother liquor is returned to the scrubber. Makeup MgO slurry is added to the mother liquor stream. Normally about 50 percent of the slurry slipstream are centrifuged. An 85 percent solids centrifuge cake, containing MgO, $MgSO_4 \cdot 7H_2O$, $MgSO_3 \cdot 6H_2O$, and $MgSO_3 \cdot 3H_2O$, is passed to a rotary dryer to dehydrate the crystals.

Dryer System

$$MgSO_4 \cdot 7H_2O \rightarrow MgSO_4 + 7H_2O$$

$$MgSO_3 \cdot 6H_2O \rightarrow MgSO_3 + 6H_2O$$

$$MgSO_3 \cdot 3H_2O \rightarrow MgSO_3 + 3H_2O$$

The dryer is direct fired and operates at about 700°F. The anhydrous solids are conveyed to a silo for storage. They will be trucked to the separate regeneration-acid plant.

At the regeneration facility, the dry product, containing about 85 percent $MgSO_3$ and 15 percent $MgSO_4$, is fed to a direct-fired rotary calciner (see Figure B-5). At 1700°F the $MgSO_3$ is converted to MgO and SO_2. Crushed green petroleum coke is added to reduce the $MgSO_4$ to MgO.

Calciner

$$MgSO_3 \rightarrow MgO + SO_2$$

$$MgSO_4 + \frac{1}{2}C \rightarrow MgO + SO_2 + \frac{1}{2}CO_2$$

The 15 percent SO_2 gas stream is used for the production of H_2SO_4. The regenerated MgO is returned to the scrubber facility as makeup.

Removal Efficiencies

The MgO process can remove 90 percent of the SO_2. Coupled with an ESP, it can reduce the particle emission by greater than 99 percent.

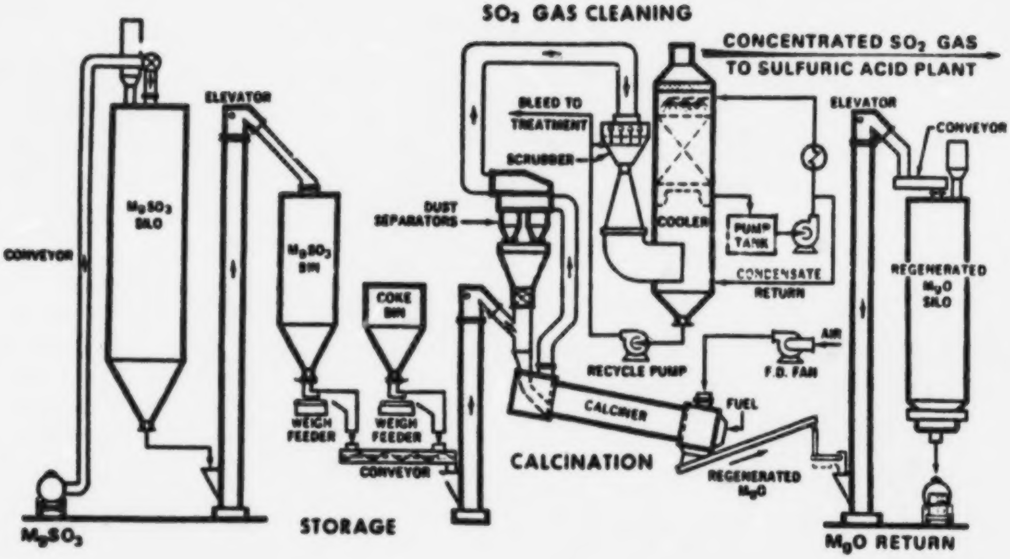

SO$_2$ GAS CLEANING
CONCENTRATED SO$_2$ GAS
TO SULFURIC ACID PLANT
ELEVATOR
BLEED TO TREATMENT
SCRUBBER
ELEVATOR
CONVEYOR
M$_g$SO$_3$ SILO
DUST SEPARATORS
COOLER
PUMP TANK
CONVEYOR
M$_g$SO$_3$ BIN
REGENERATED M$_g$O SILO
CONDENSATE RETURN
COKE BIN
AIR
RECYCLE PUMP
F.D. FAN
FUEL
CALCINER
WEIGH FEEDER
WEIGH FEEDER
CALCINATION
CONVEYOR
REGENERATED M$_g$O
STORAGE
M$_g$SO$_3$
M$_g$O RETURN

FIGURE B-5. MgO REGENERATION PROCESS FLOW SHEET(60)

Wastes

The only waste stream is the fly ash slurry.

Raw Material and Heat Requirements

Makeup magnesium oxide and coke are required in the scrubbing and regeneration processes. If an acid plant is included in the overall operation, makeup catalyst also is necessary. Fuel oil is required to heat the dryer and calciner.

By-Product

The regenerated SO_2 gas is utilized in the manufacture of H_2SO_4. About 1.33 lb of sulfuric acid (98 percent) is produced per pound of SO_2 removed.

Advantages

(1) High SO_2 removal efficiency, 90 percent
(2) Absorption and regeneration steps can be separated
(3) Minimal solid wastes disposal problem.

Disadvantages

(1) Stack gas reheat may be required
(2) High energy requirements for drying and calcining
(3) High cost of MgO absorbent
(4) Absorption facility must be located near a regeneration facility.

Development Status

After several pilot plant studies, a commercial size system was constructed on Boston Edison's oil-fired 155-MW boiler in 1971. It operated on and off for 2 years. The spent magnesium salt was transported to the Essex Chemical Company's acid plant in Rumford, Rhode Island, for regeneration.

To gain experience with a coal-fired facility, an MgO scrubbing system was constructed at Potomac Electric Power Company's (PEPCO's) Dickerson Unit 3, 190-MW boiler to treat half the flue gas, 295,000 acfm. The system was placed in operation in September, 1973. Initial operation, debugging, and modifications have been made since then and the system is now being analyzed for optimum operation. The SO_2 removal efficiency has been in excess of 90 percent. PEPCO's estimate for the capital cost is in excess of $100/kW in 1974 dollars. The operating cost is not available due to lack of pertinent data.

Capital and Operating Cost

TVA(61) estimated the capital cost of an MgO system for a 200-MW existing boiler burning 3.5 percent sulfur coal with on-site regeneration facilities at $\$13.1 \times 10^6$ in 1972. This included $\$4.58 \times 10^6$ for the calcining and sulfuric acid manufacturing plants. The labor, material, and utility requirements are shown in Table B-6.

Shah and Quigley(62) estimated the labor, material, and utility requirements for an MgO system excluding the regeneration facilities and the results are shown in Table B-7. They also estimated the capital cost for a central regeneration station including calcination and sulfuric acid plants (capacity, 1000 tpd of sulfuric acid) at $\$8.2 \times 10^6$ in 1972. Table B-8 shows the labor, material, and utility requirements of the regeneration station.

TABLE B-6. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR INTEGRATED MgO SYSTEM(61)

Basis: 200 MW existing boiler
Coal fired, 554,200 tons/yr
3.5 percent sulfur in coal
Operation, 7,000 hours/yr
H_2SO_4, 46,600 tons/yr

Item	Quantity Coal fired
Utility	
Power, MW	4
Fuel oil, gal/hr	452
Process water, gpm	2,200
Material	
MgO, tons/hr	0.07
Coke, tons/hr	0.05
Labor	
Direct operation, men/shift	2
Maintenance	7 percent of capital cost

TABLE B-7. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR MgO SCRUBBING PROCESS (excluding calcination and acid production)(62)

Basis: 600-MW oil-fired boiler
2.5 percent sulfur oil
Load factor, 65 percent
Removal efficiency, 90 percent
Fuel consumption, 4,500,000 bbl/yr

Item	Quantity
Utility	
Power	5.4 MW
Fuel oil	14.3 bbl/hr
Water	430 gpm
Material	
Makeup MgO	0.074 tons/hr
Labor	
Direct operation	2.3 men/shift
Maintenance	4 percent of capital cost/yr

TABLE B-8. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR CENTRAL REGENERATION AND ACID PRODUCTION PLANT(62)

Basis: Capacity, 1000 tons/day of H_2SO_4
Load factor, 100 percent
Operation, 330 days/yr

Item	Quantity
Utility	
Power	2 MW
Boiler feed water	88 gpm
Process water	28 gpm
Cooling water	417 gpm
Fuel oil	31.7 bbl/hr
Material	
Coke	0.1 tons/hr[a]
Labor	
Direct operation	8.3 men/shift
Maintenance	4 percent of capital cost

[a] This value seems small compared with that obtained from TVA study.(61)

Wellman-Lord Process

Developer/Manufacturer

The Wellman-Lord SO_2 removal process was developed by Davy Powergas in the late 1960's. The process is marketed internationally by Davy Powergas.

Process Description

The Wellman-Lord process employs a sodium sulfite scrubbing solution to remove SO_2 from flue gas. Thermal regeneration is utilized to recover the sulfite and produce a by-product stream of 95 percent SO_2.

Flue gas, at approximately 270°F is compressed by a booster fan (see Figure B-6) and passed through a venturi scrubber to remove the fly ash. The scrubbing liquid, containing the fly ash, flows into a recirculation sump from where it is withdrawn and recycled to the venturi. A purge stream is continuously withdrawn from the sump and transported to the fly ash solids handling area. The flue gas proceeds into a fire stage absorber where it contacts a countercurrent 27 percent solution of Na_2SO_3, $NaHSO_3$, and Na_2SO_4.

The SO_2 reacts with $SO_3^=$ to form HSO_3^-; some oxidation of $SO_3^=$ to $SO_4^=$ occurs.

$$SO_2 + SO_3^= + H_2O \rightarrow 2HSO_3^-$$

$$SO_3^= + 1/2\ O_2 \rightarrow SO_4^=$$

The purified gas flows through a chevron mist eliminator, out of the absorber, and to the stack. The flue gas, at 125°F, is generally not reheated.

The spent scrubber solution flows from the bottom stage of the absorber and separates into two streams. One stream, about 10 percent of the total flow is sent to the purge treatment system for sodium sulfate removal, the remainder is sent to a surge tank. From the surge tank, the absorber slurry is heated in a heat exchanger and introduced to a

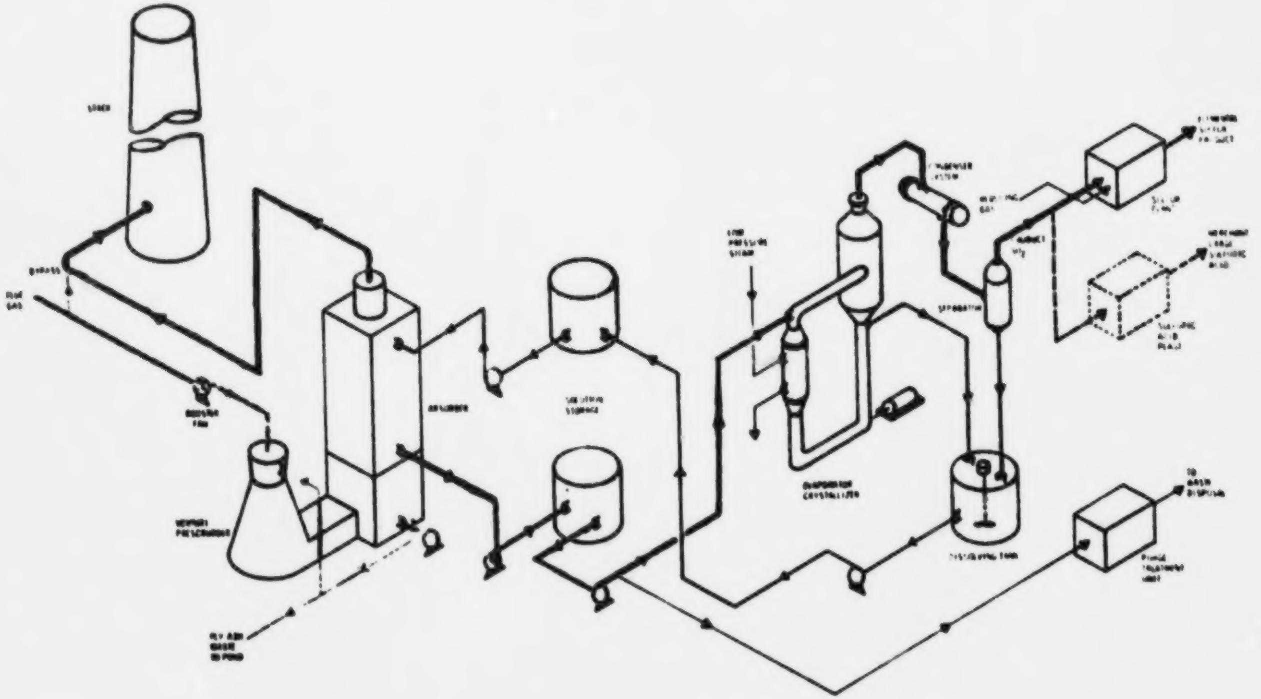

STACK
BYPASS
FLUE GAS
BOOSTER FAN
VENTURI PRESCRUBBER
ABSORBER
FLY ASH WASTE TO POND
SOLUTION STORAGE
LOW PRESSURE STEAM
EVAPORATOR CRYSTALLIZER
CONDENSER SYSTEM
SEPARATOR
REDUCING GAS
PRODUCT SO2
SULFUR PLANT
ELEMENTAL SULFUR PRODUCT
SULFURIC ACID PLANT
DISSOLVING TANK
PURGE TREATMENT UNIT
TO WASTE DISPOSAL

FIGURE B-6. WELLMAN-LORD SO_2 RECOVERY PROCESS(63) double-effect evaporator. Sodium bisulfite decomposes to sodium sulfite releasing water and sulfur dioxide:

$$2HSO_3 \rightarrow SO_3 + SO_2 + H_2O \quad .$$

A disproportionation reaction takes place in the evaporator at high temperatures:

$$6Na^+ + 6HSO_3^- \rightarrow 2Na_2SO_4 + Na_2S_2O_3 + 2SO_2 + 3H_2O \quad .$$

Sodium sulfate and thiosulfate ($Na_2S_2O_3$) formed in this reaction are removed from the solution in the purge treatment system.

The overhead vapors, containing SO_2 and H_2O, are cooled and sent to the SO_2 stripper. The stripper overhead vapor is cooled to reduce the water content to 5 percent by weight. It is then heated, compressed and sent to an acid or sulfur processing plant. Condensate from the stripper is used to slurry the Na_2SO_3 crystals in the dissolving tank. Either NaOH or Na_2CO_3 is added to the dissolving tank to make up for the sodium lost in the purge stream.

The purge stream is first cooled in a heat exchanger followed by additional cooling in a chilled vessel. It flows to an ethylene glycol refrigerated crystallizer where the less soluble Na_2SO_4 precipitates. The crystallizer bottoms are transferred to a thickener. The thickener underflow is sent to a centrifuge, and the thickener overflow is returned to the crystallizer.

Wet cake from the centrifuge drops into a jacketed dryer where any sodium pyrosulfite decomposes to sodium sulfite and SO_2:

$$Na_2S_2O_5 \rightarrow Na_2SO_3 + SO_2$$

SO_2 vapor is vented to the flue gas handling system. The dried cake is transported to a storage bin.

Mother liquor from the centrifuge flows to a purge tank and is returned to the absorber product surge tank. Normally about 50 percent of the Na_2SO_4 formed during aboorption and regeneration is removed in the purge treatment system.

Removal Efficiencies

The Wellman-Lord system is capable of removing 90 percent of the SO_2 and 99 percent of the fly ash from flue gas. This system has no NO_x removal capability.

Raw Material and Heat Requirements

Raw materials required in Wellman-Lord process include soda ash and water. Steam is required for the evaporator operation.

Materials of Construction

The absorber is made of a tile-lined carbon steel with 316 L stainless steel internals. All flow lines and surge tanks are rubber lined. The slurry pumps, heat exchangers, evaporator, stripper, and purge system process equipment are 316 L stainless steel.

By-Product

A 95 percent SO_2 off-gas stream is produced. It may be utilized in the production of sulfuric acid or elemental sulfur.

Wastes

There are two waste streams, the fly ash slurry from the venturi and the purge stream. The rate of purging is estimated to be 0.6 lb solids content (32.5 percent)/lb of sulfur removed.

Advantages

(1) Small waste generation
(2) High particle and SO_2 removal efficiency

(3) Proven reliability and performance in commercial installations (for oil-fired boiler only)

(4) No plugging or scaling in scrubber

(5) Low liquid to gas ratio, normally about 10 gal/10^3ft^3.

Disadvantages

(1) High heat and energy requirements

(2) Complicated process

(3) SO_2 processing plant required

(4) Not a compact system

(5) Stack gas reheat may be required

(6) Not demonstrated on coal-fired boiler.

Development Status

This process has been applied almost exclusively to acid plant tail gases and oil-fired boilers as shown in Tables B-9 and B-10. The first commercial coal-fired power plant installations are under construction at Northern Indiana Public Service Company's 115-MW boiler in Gary, Indiana, and Public Service Company of New Mexico's 700-MW boiler in Fruitland, New Mexico.

Capital and Operating Costs

The mid-1974 capital cost estimate for a 500-MW coal-fired boiler with facilities to reduce the off-gas SO_2 to elemental sulfur is \$33 x 10^6.(63) The cost for the same unit with an acid plant is \$30 x 10^6.(63) The labor, material, and utility requirements are shown in Table B-11.

TABLE B-9. COMMERCIAL APPLICATIONS OF WELLMAN-LORD PROCESS(63)

Company and Location	Start-Up Date	Type and Size of Plant	Operational Time, years
Olin Corporation Paulsboro, NJ	July, 1970	Sulfuric Acid Plant 700 tpd	4.0
Toa Nenryo Kawasaki, Japan	August, 1971	Sulfur Recovery Plants 2 @ 150 tpd each	3.0
Japan Synthetic Rubber Co. Chiba, Japan	August, 1971	2 Oil Fired Boilers (70 MW equivalent)	3.0
Standard Oil of California El Segundo, CA	September, 1972	Sulfur Recovery Plants 3 @ 150 tpd each	1.11
Allied Chemical Corporation Calumet, IL	November, 1972	3 Sulfuric Acid Plants Total cap. 500 tpd	1.9
Olin Corporation Curtis Bay, MD	May, 1973	3 Sulfuric Acid Plants Total cap. 1000 tpd	1.2
Sumitomo Chiba Chemical Co. Sodegaura, Japan	November, 1973	Oil-Fired Boiler (125 MW equivalent)	0.9
Japanese Synthetic Rubbe- Yokkaichi, Japan	December, 1973	Oil-Fired Boiler (140 MW equivalent)	0.8
Kashima Oil Company Kashima, Japan	February, 1974	Sulfur Recovery Plants 2 @ 90 tpd each	0.7
Chubu Electric Company Nagoya, Japan	May, 1973	Oil-Fired Boiler 220-MW power plant	1.4

TABLE B-10. WELLMAN-LORD SO_2 RECOVERY SYSTEMS EITHER UNDER DESIGN OR CONSTRUCTION

Company and Location	Type of Plant	Flue Gas Rate Through Units, SCF
Confidential Client Kawasaki, Japan	Steam boiler	435,000
Kashima Mutual Power Kashima, Japan	Oil-fired power plant	590,000
Kuraray Okayama, Japan	Oil-fired boiler	248,000
Mitsubishi Chemical Mitzushima, Japan	Oil-fired boiler	373,000
Northern Indiana Public Service Co. Gary, IN	Coal-fired power plant	310,000
Public Service Co. of New Mexico Fruitland, NM	Coal-fired power plant	1,800,000
Standard Oil Co. of California Richmond, CA	Claus plants	30,000
Standard Oil Co. of California El Segundo, CA	Claus plant	30,000
Toa Nenryo Kogyo K.K. Hatsushima, Japan	Claus plant	10,000
Toyo Rayon Nagoya, Japan	Oil-fired boiler	218,000
Toa Nenryo Kogyo K.K. Arita, Japan	Claus plant	34,000

TABLE B-11. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR A WELLMAN-LORD PROCESS

Basis: 500-MW boiler
3.5 percent sulfur coal
100 percent load factor, 330 days/yr

Item	Quantity	
	Elemental Sulfur	Sulfuric Acid
Utilities		
Power, MW	16.0	16.4
Steam, lb/hr	1.68×10^5	1.88×10^5
Water, gpm	4,900	7,000
Material		
Soda ash, tons/hr	0.66	0.66
Natural gas, 10^3 scf/hr	114	--
Labor		
Direct operation, man/shift	5	5
Maintenance	4 percent of capital/hr	4 percent of capital/yr

Double Alkali Process (GM)

Developer/Manufacturer

GM developed the dilute double alkali process installed at their Parma, Ohio, auto plant.

Process Description

This process employs a dilute sulfite buffered scrubbing solution **with lime regeneration. Calcium sulfate plugging is minimized** by **softening with sodium carbonate.**

First the flue gas is treated for fly ash removal and then introduced into the bottom of a scrubber where it is contacted with a slurry spray and cooled to prevent wet-dry interphase scaling (see Figure B-7). The gas flows upward through three trays where SO_2 in the gas is removed by reacting with a 0.1 molar caustic soda slurry at a liquor to gas ratio of 20 gal/10^3 lb. Each tray contains a series of floating bubble caps which rise and fall with changing gas flow. The following reactions occur.

$$2OH^- + SO_2 = SO_3^= + H_2O$$

$$SO_3^= + SO_2 + H_2O = 2HSO_3^-$$

$$SO_3^= + 1/2\ O_2 = SO_4^=$$

The normal pH of the scrubbing slurry is 5.0. System pressure drop is about 7.5 inches water. The clean gas passes through a mesh mist eliminator and is vented to atmosphere. Scrubbing slurry is pumped from the bottom of the absorber where 75 percent is recycled to the top of the absorber and the remainder is pumped to a mix tank. In the first mix tank $CaCO_3$ converts bisulfite to sulfite and $CaSO_3$ precipitates.

$$HSO_3^- + CaCO_3 = CaSO_3 + CO_2 + SO_3^= + H_2O$$

In the second mix tank, lime is added to causticize Na_2SO_3 and Na_2SO_4.

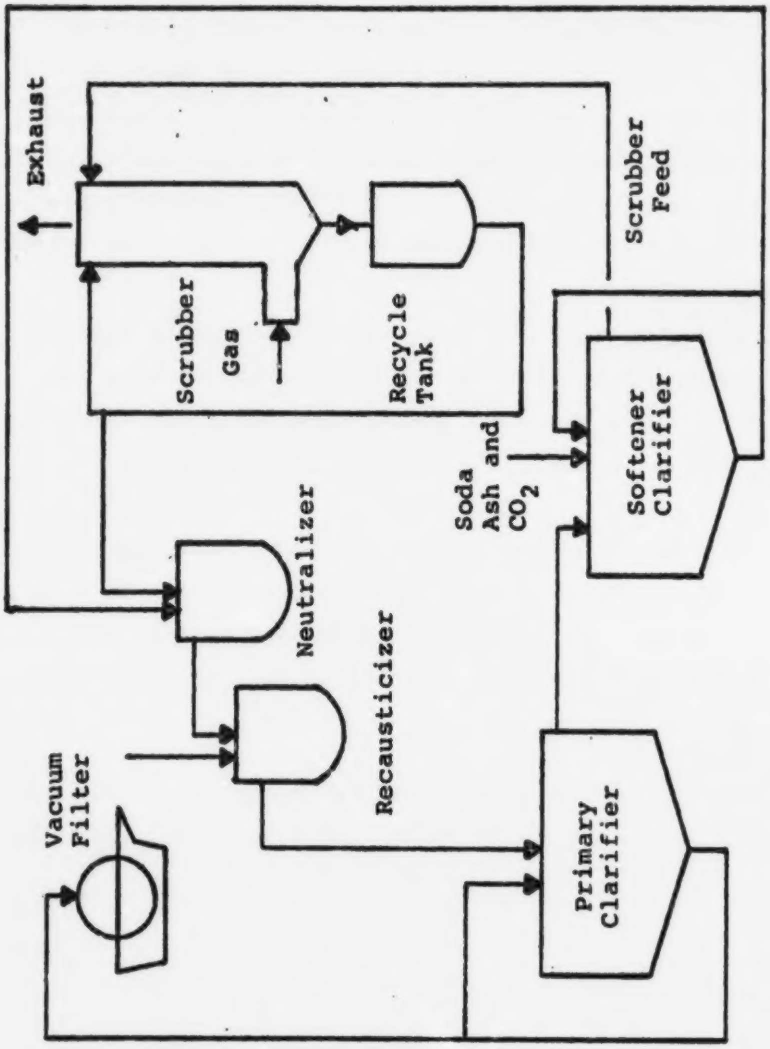

Vacuum Filter
Recausticizer
Neutralizer
Primary Clarifier
Softener Clarifier
Soda Ash and CO_2
Scrubber
Gas
Recycle Tank
Exhaust
Scrubber Feed

FIGURE B-7. GM DOUBLE ALKALI PROCESS FLOW SHEET(87)

$$Ca(OH)_2 + SO_3^{=} = 2OH^- + CaSO_3$$

$$Ca(OH)_2 + SO_4^{=} = 2OH^- + CaSO_4$$

Because of the formation of Na_2SO_4, the regeneration section has to be capable of regenerating both sulfate and sulfite. The sulfate reaction, however, is quite difficult because of the relative solubility of the product, calcium sulfate. In addition, the sodium sulfate cannot be causticized in the presence of high $SO_3^{=}$ or OH^- because Ca^{++} levels are held below $CaSO_4$ solubility product. Thus, to provide effective sulfate regeneration, the system must be operated at a dilute OH^- concentration (<0.14M) while maintaining sufficient levels of sulfate (>0.4M) to effect calcium sulfate precipitation.

The solution is pumped to the reactor-clarifier where the calcium salts precipitate out of solution. The clarifier underflow is dewatered in a vacuum filter to about 50 percent moisture. The filter cake is high in $CaSO_4$ with some $CaSO_3$, fly ash, and small amounts of sodium contaminants.

Clarifier overflow, containing about 800 ppm of calcium ion, and saturation amounts of sulfate, sulfite, and hydroxide ions, is pumped to a second reactor clarifier. Soda ash is introduced to soften the solution by precipitating $CaCO_3$ which is recycled to the first mix tank.

$$CO_3^{=} + Ca(OH)_2 = 2OH^- + CaCO_3$$

The soda ash also serves to replace sodium lost to the filter cake. The regenerated slurry, containing about 250 ppm of calcium, is recycled to the scrubber.

Removal Efficiencies

The SO_2 removal efficiency varies between 88 and 92 percent. The system is capable of reducing particles from an inlet loading of 0.3 gr/scf to 0.05 gr/scf.

Materials of Construction

The scrubbers are constructed of 316L stainless steel.

Wastes

The only waste stream is the filter cake. It contains primarily $CaSO_4$ with fly ash, $CaSO_3$, and some sodium salts.

By-Product

None.

Raw Materials and Heat Requirements

As GM does not reheat the flue gas, no heat is required for the process. Soda ash and lime are consumed as raw materials.

Advantages

There is insufficient information on the operating performance of this process to quantitatively appraise it. Listed below are the conclusions drawn from information obtained from GM performance reports and other similar processes.

(1) High SO_2 removal efficiency
(2) No severe plugging or scaling problems.

Disadvantages

(1) High oxidation rate results in difficulty in controlling pH of incoming slurry
(2) Waste sludge disposal requirement
(3) High lime and soda ash cost
(4) Dilute OH^- ion concentration requires circulating large quantities of slurry.

Development Status

Following a pilot plant operation in 1969, GM completed a full-scale double alkali SO_2 removal facility on a 400,000 lb steam/hour capacity coal-fired boiler in Parma, Ohio. It was started up in March, 1974 and a 1-year in-depth evaluation of the total system is now in progress. GM intends to apply a similar system to GM's other industrial boilers.

Capital and Operating Costs

GM estimated the cost of the double alkali system for a coal-fired boiler equivalent to 32 MW at $\$3.5 \times 10^6$ in mid-1973.(87) The labor, material, and utility requirements were estimated as shown in Table B-12.

TABLE B-12. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR GM DOUBLE ALKALI PROCESS(87)

Item	Quantity[b]
Utility	
Power	400 kW
Steam	2,700 lb/hr
Water	21.4 gpm
Material	
Lime	0.23 tons/hr
Soda ash	0.05 tons/hr
Carbon dioxide	0.007 tons/hr
Polymer	0.03 lb/hr
Supplies	1 percent of capital cost/yr[a]
Labor	
Direct operation	1.4 men/shift
Maintenance	0.5 man/shift

(a) This value was assumed.

(b) The operating load factor was 0.47.

Chemiebau Process

Developer/Manufacturer

The Chemiebau process was developed by Reinluft GmbH, Essen, Germany, and acquired by Chemiebau - Dr. A. Zieren, GmbH, Cologn, Germany, in 1967. Commonwealth Associates, Jackson, Michigan, has acquired the western hemisphere licensing rights.

Process Description

The process employs moving beds of lump char to remove SO_2 from flue gas.(88) The absorbent is thermally regenerated, producing a concentrated 20 percent SO_2 gas stream. A schematic process flow sheet is shown in Figure B-8.

The flue gas, following treatment by a mechanical dust collector, is introduced into the bottom of the adsorber. The adsorber is a steel shell containing vertical shafts through which walnut-size lumps of charcoal flow. Baffles and louvers are installed to direct the gas cross countercurrent against the downward moving beds of char. At an optimum operating temperature of 250 to 300°F, the SO_2 is adsorbed by the dry char and catalytically converted to SO_3. The SO_3 reacts with water to form H_2SO_4 which condenses within the char.

$$SO_2 + 1/2\ O_2 \rightarrow SO_3$$

$$SO_3 + H_2O \rightarrow H_2SO_4$$

The purified gas exits from the top of the adsorber, passes through an ID fan and is vented to the stack.

Conveyor belts and bucket elevators transport the acid-laden char from the bottom of the adsorber into the top of the desorber. The desorber unit is similar in structure to the adsorber; char flows down through vertical shafts where it is heated to about 700°F by a cross-countercurrent flow of inert scavenger gas. Sulfuric acid dissociates into SO_3 and water, and the SO_3 reacts with carbon to form SO_2 and CO_2.

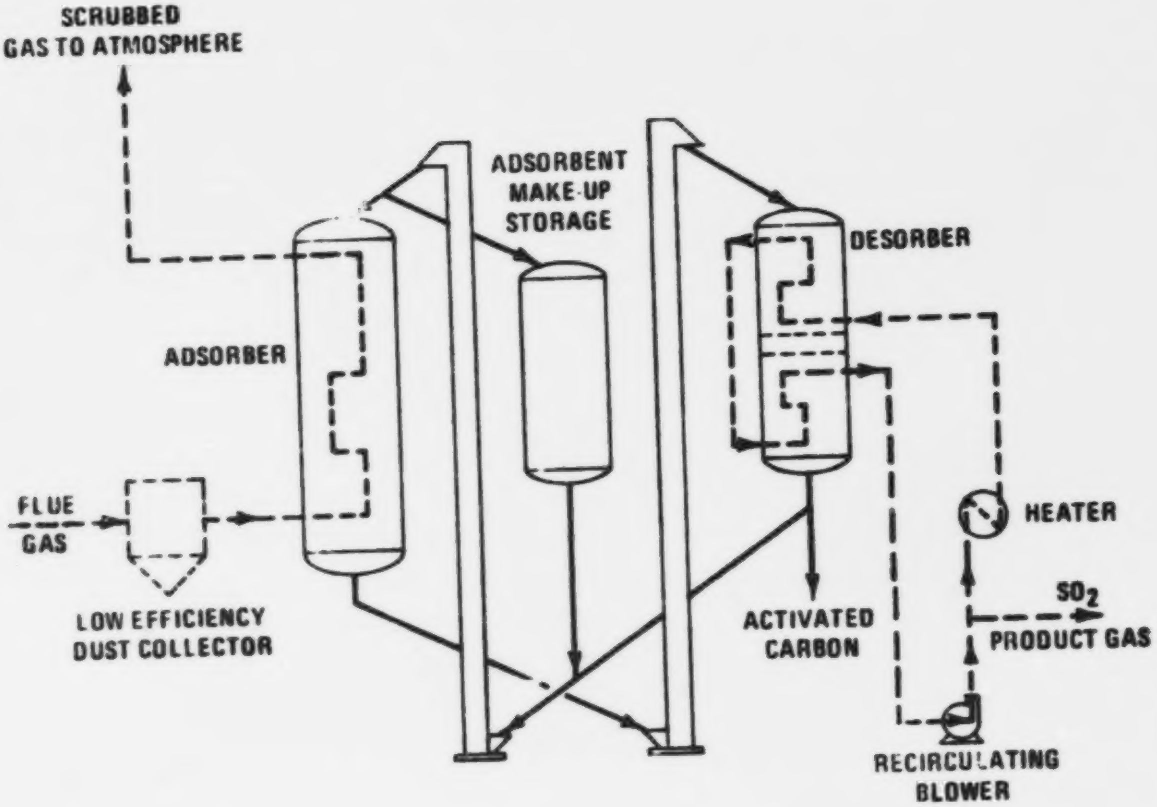

SCRUBBED GAS TO ATMOSPHERE
ADSORBENT MAKE-UP STORAGE
DESORBER
ADSORBER
FLUE GAS
HEATER
SO_2 PRODUCT GAS
LOW EFFICIENCY DUST COLLECTOR
ACTIVATED CARBON
RECIRCULATING BLOWER

FIGURE B-8. CHEMIEBAU FLOW SHEET(88)

$$H_2SO_4 \rightarrow H_2O + SO_3$$

$$2SO_3 + C \rightarrow CO_2 + 2SO_2$$

The scavenging gas, containing about 20 percent SO_2, passes from the desorber into a circulating fan. At the fan discharge, a portion of the gas is withdrawn to a sulfur recovery system at a rate equivalent to the amount of SO_2 generated in the desorber. The SO_2 product recovery stream is reheated and returned to the desorber. Regenerated adsorbent is discharged from the bottom of the desorber, screened to remove the fine particles, and returned to the adsorber. Makeup char, stored in the char bin, is added to the adsorber to replace the char lost due to mechanical attrition and reaction with SO_3.

Removal Efficiencies

The Chemiebau system is capable of a 95 percent SO_2 removal efficiency. Fly ash can also be removed by the system and has no adverse effect on the adsorbent activity. Although the specific information is not available, the system has appreciable NO_x and halogen removal capability.

Wastes

There are no waste streams.

Materials of Construction

The adsorber and desorber are constructed of carbon steel. The conveyor system from the desorber must be constructed of materials capable of handling the hot regenerated char particles.

By-Product

In addition to the 20 percent SO_2 off-stream produced, the char removed in the screening process is very active and can be marketed as activated carbon. About 0.[illegible] [illegible] of activated carbon would result per pound of SO_2 removed.

Raw Material and Heat Requirements

Adsorbent char is required in the adsorber at a rate of about 20 lb per pound of SO_2 removed. About 4 percent of the char would be lost per cycle due to mechanical attrition and reaction with SO_3, and, thus, the makeup char must be added for the loss. Low cost lignite serves as the adsorbent char.

Direct fired heat is required to reheat the scavenger gas from 500° to 750° F for adsorbent regeneration. Assuming an 80 percent heater efficiency, the heat requirement would be 5,000-6,000 Btu/lb of SO_2 removed.

Advantages

(1) No flue gas reheat
(2) Capable of removing NO_x and halogens
(3) No waste streams
(4) Good turndown capability
(5) Possible credit for activated by-product.

Disadvantages

(1) High char attrition rate
(2) Low space velocities require massive adsorber and desorber vessels
(3) Requirement of concentrated SO_2 processing system.

Disadvantages

(1) High char attrition rate

(2) Low space velocities require massive adsorber and desorber vessels

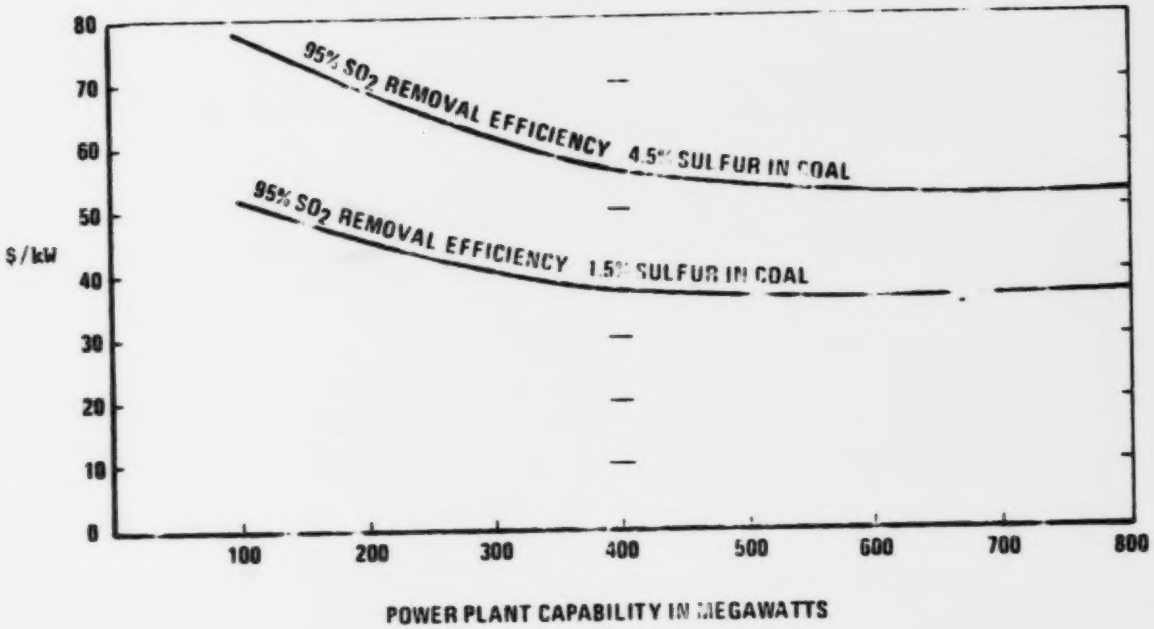
(3) Requirement of concentrated SO_2 processing system.

Development Status

The most recent installation was a 10-MW pilot plant test facility at Kellerman Power Station, Lunen, Germany, 1966-68. Commonwealth Associates, Jackson, Michigan, is the Chemiebau licenser for the Western Hemisphere. To date no Chemibau processes have been sold in the United States.

Capital and Operating Costs

Figure B-9 shows the estimated capital costs for the Chemiebau system in November, 1973.(88) The estimates were based on field erection costs of two 50 percent capacity trains ready to run with initial loading of adsorbents. The cost included the mechanical conveying equipment, a char storage bin, an 80 percent efficiency mechanical dust collector, and the equipment to reduce the SO_2 to elemental sulfur. The operating labor, material, and utility requirements for a Chemiebau process including elemental sulfur production system installed on a 100 MW equivalent boiler are shown in Table B-13. A credit may be taken into consideration for a by-product activated carbon produced at a rate of 2.16 tons/hr.



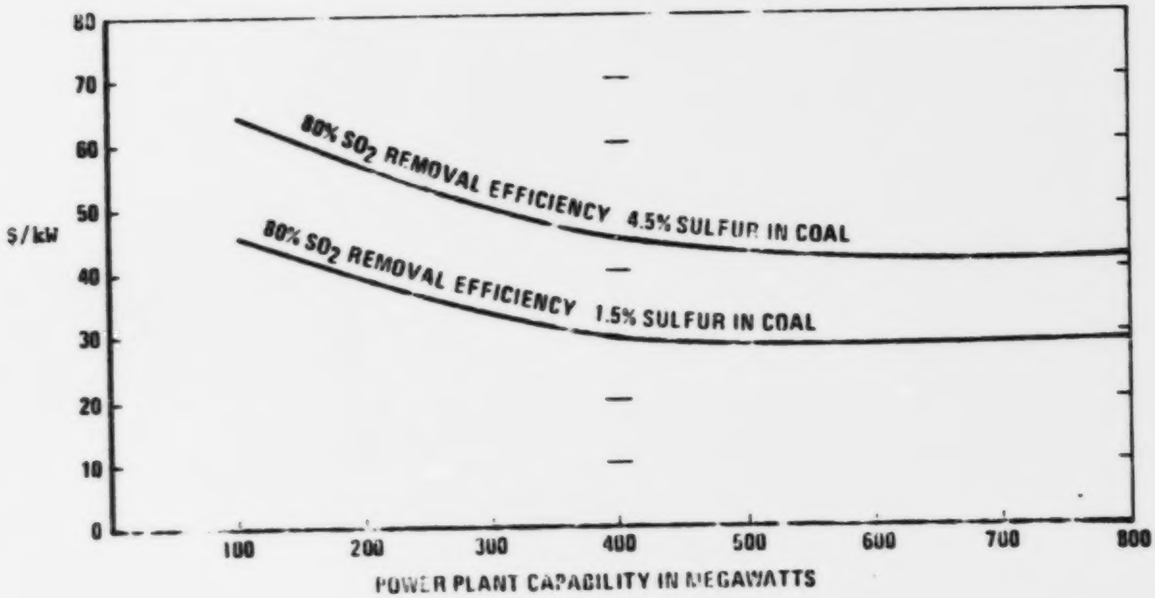


FIGURE B-9. CAPITAL COST FOR CHEMIEBAU PROCESS - INCLUDING 30 DAYS CHAR SUPPLY AND SULFUR REDUCTION - TWO 50 PERCENT TRAINS AT A NEW INSTALLATION(88)

TABLE B-13. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR CHEMIEBAU PROCESS(89)

Basis: Coal-fired boiler (100 MW equivalent)
4.5 percent sulfur in coal (SO_2 conc., 3,600 ppm)
95 percent removal efficiency
85 percent load factor
Heating value, 12,000 Btu/lb

Item	Quantity
Utility	
Power	1.3 MW
Fuel oil	1,850 lb/hr
Raw Material	
Lignite	2.99 tons/hr
Labor	
Direct operation	1.5 men/shift
Maintenance	4.5 percent of capital cost

Foster Wheeler (FW)

Developer/Manufacturer

The FW process for SO_2 removal is a combination of char adsorption and regeneration processes developed by Bergbau Forschung, GmbH, and elemental sulfur conversion process developed by Foster Wheeler. Foster Wheeler currently markets the process in the United States.

Process Description

After treatment for removal of particulate matter, the flue gas, at about 300°F, is introduced into the adsorber (see Figure B-10). The adsorber contains vertical parallel louver beds through which the char flows. The char moves in a plug flow fashion with the flow rate controlled by a vibratory feeder located at the bottom of each bed, and the flue gas passes through the adsorber bed in a cross flow. A portion of SO_2 adsorbed in the char is converted to H_2SO_4 by the following reaction.

$$SO_2 + 1/2\ O_2 \rightarrow SO_3$$

$$SO_3 + H_2O \rightarrow H_2SO_4$$

The acid laden char pellets flow from the bottom of the adsorber where they are screened for fly ash and are conveyed to the top of the regeneration unit.

In the regenerator the char is heated by mixing with hot sand at 1500°F. The sand serves as an inert heat transfer media. The following reactions take place.

$$H_2SO_4 \rightarrow H_2O + SO_3$$

$$2SO_3 + C \rightarrow 2SO_2 + CO_2 \quad .$$

SO_2 gas is liberated and the regenerated char pellets are discharged from the bottom of the vessel to the separator where they are separated from the sand, cooled, and recycled to the adsorber.

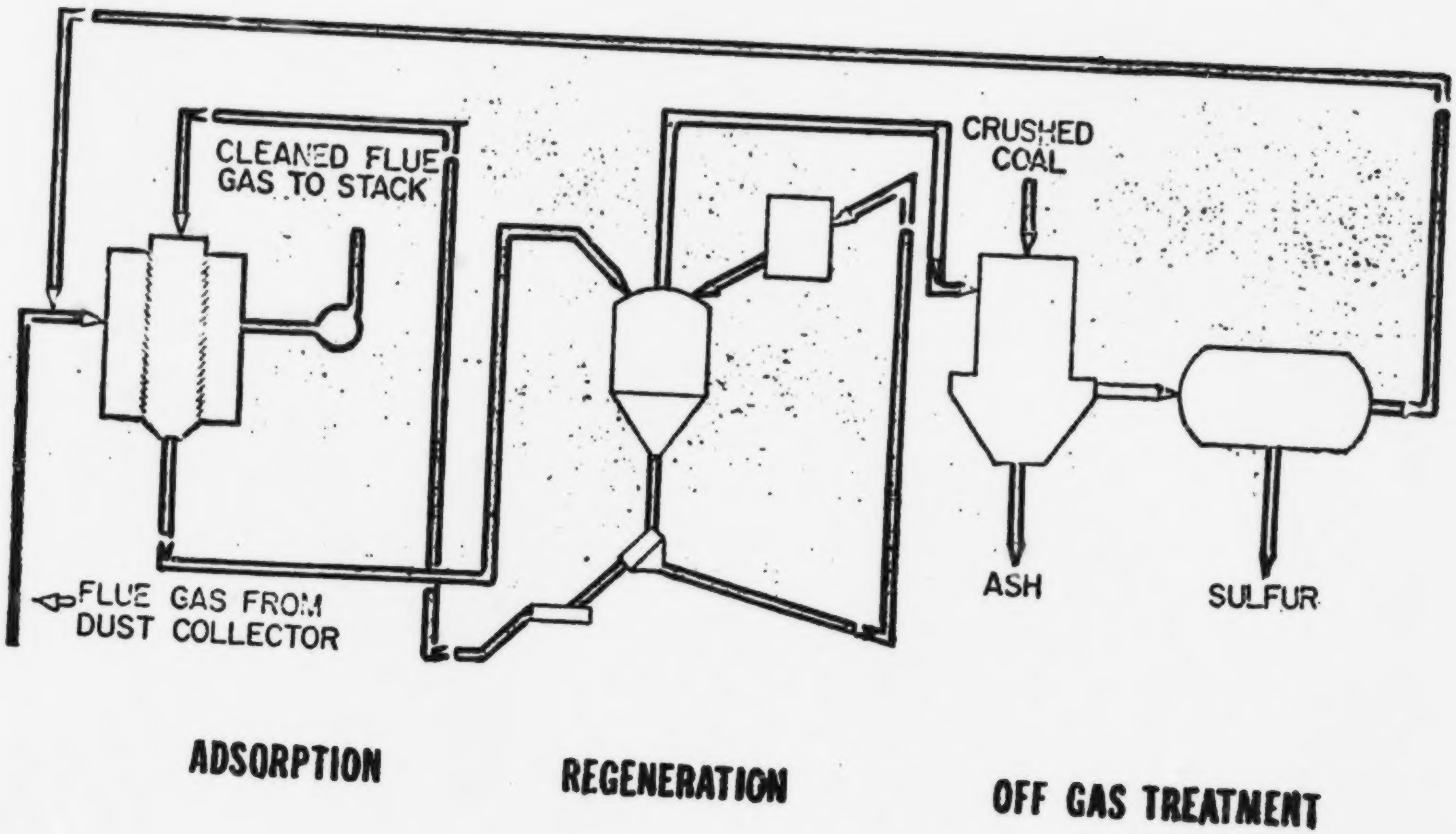

CLEANED FLUE GAS TO STACK
CRUSHED COAL
FLUE GAS FROM DUST COLLECTOR
ASH
SULFUR
ADSORPTION
REGENERATION
OFF GAS TREATMENT

FIGURE B-10. PROCESS FLOW OF FW SYSTEM(90)

The concentrated SO_2 gas stream is directed to the Foster Wheeler off-gas treatment system where SO_2 reacts with crushed coal and is reduced to elemental sulfur.

$$C + SO_2 \rightarrow CO_2 + S$$

The resulting gases enter a condenser where the sulfur is condensed and stored in a heated tank. The remaining gases are recycled to the adsorber to capture any remaining SO_2.

Removal Efficiencies

The system is capable of removing 86-95 percent of the SO_2, 90-95 percent of the particulate matter, and 40-60 percent of the NO_x.

Raw Materials and Heat Requirements

About 0.14 lb char is lost per pound of SO_2 adsorbed. These losses are primarily due to the production of CO_2 in the regenerator (about 90 percent) with the remainder (about 10 percent) attributed to mechanical attrition. Twice annually, the entire char system is replaced by new char.

Heat is required to heat the sand to 1500°F for the regeneration process.

Wastes

None.

Advantages

(1) No flue gas reheat
(2) Production of marketable sulfur by-product
(3) Good turndown capability
(4) Capable of removing NO_x

(5) No waste streams

(6) Coal is the reducing agent for sulfur production.

Disadvantages

(1) Char must be replaced twice annually

(2) Solids handling equipment presents possible maintenance problems

(3) Low space velocities require massive adsorber and regenerator vessels.

By-Product

About 0.5 lb of elemental sulfur is produced per pound of SO_2 removed.

Development Status

Bergbau-Forschung started up a demonstration unit in early 1974 at the Kellerman Power Plant in Lunder, West Germany. The unit processes flue gas equivalent to about 35 MW as a slip stream from a 350-MW coal-fired boiler. The off gas is treated in a Claus reactor. Foster Wheeler is installing a demonstration unit at Gulf Power Company's Scholz Steam Plant on a 47.5-MW coal-fired boiler. The adsorption section is designed for half load (50 percent of flue gas flow) and the regeneration section is designed for full load so that higher sulfur coal can be tested. The 1-year test program is under way.

Capital and Operating Costs

The capital cost for a turnkey installation on a 500-MW boiler system, burning 3 percent sulfur coal was estimated at $55 to $70/kW in mid-1974.(91) The labor, material, and utility requirements for a 20-MW equivalent system are shown in Table B-14.(91)

TABLE B-14. LABOR, MATERIAL, AND UTILITY REQURIEMENTS FOR FW PROCESS(91)

Basis: 20-MW coal-fired boiler
3 percent sulfur in coal
100 percent load factor
Heating value, 12,000 Btu/lb
90 percent removal efficiency

Item	Quantity
Utility	
Power	500 kW
Fuel oil	500 lb/hr
Material	
Makeup char	126 lb/hr
Char	
Labor	
Direct operation	1 man/shift
Maintenance	4 percent of capital cost

Westvaco Process

Developer/Manufacturer

Westvaco, Charleston Heights, South Carolina, developed and markets the Westvaco SO_2 process.

Process Description

The Westvaco SO_2 process uses activated carbon to adsorb dilute SO_2 from flue gas. Upon regeneration of the carbon, H_2S is utilized to reduce the entrained H_2SO_4 to elemental sulfur.

Following treatment for particle removal, the flue gas is introduced to a five-stage activated carbon fluidized-bed adsorption unit (see Figure B-11). SO_2 is removed through catalyzed oxidation to SO_3 and a subsequent hydrolysis to sulfuric acid which remains adsorbed in the carbon particles.

$$SO_2 + 1/2\ O_2 + H_2O \rightarrow H_2SO_4$$

Sufficient water vapor and oxygen are present normally in the flue gas for the reaction. The purified gas exits through the top of the adsorber and is vented to the stack.

The acid loaded carbon is transferred mechanically to the sulfur generator where, at about 300°F, it is contacted by a stream of H_2S. The H_2S reduces the H_2SO_4 to elemental sulfur. In general, hydrogen sulfide is not available in all industrial boilers, and therefore, the following reaction is added.

$$3H_2 + 4S \xrightarrow{\text{activated carbon}} 3H_2S + S$$

The reaction takes place in the H_2S generator/sulfur stripper at temperatures near 1000°F. The hydrogen required may be possibly supplied through a gasifier utilizing coal or other fossil fuels. The mixture of H_2S and sulfur vapor leaves the H_2S generator/sulfur stripper and passes to the sulfur

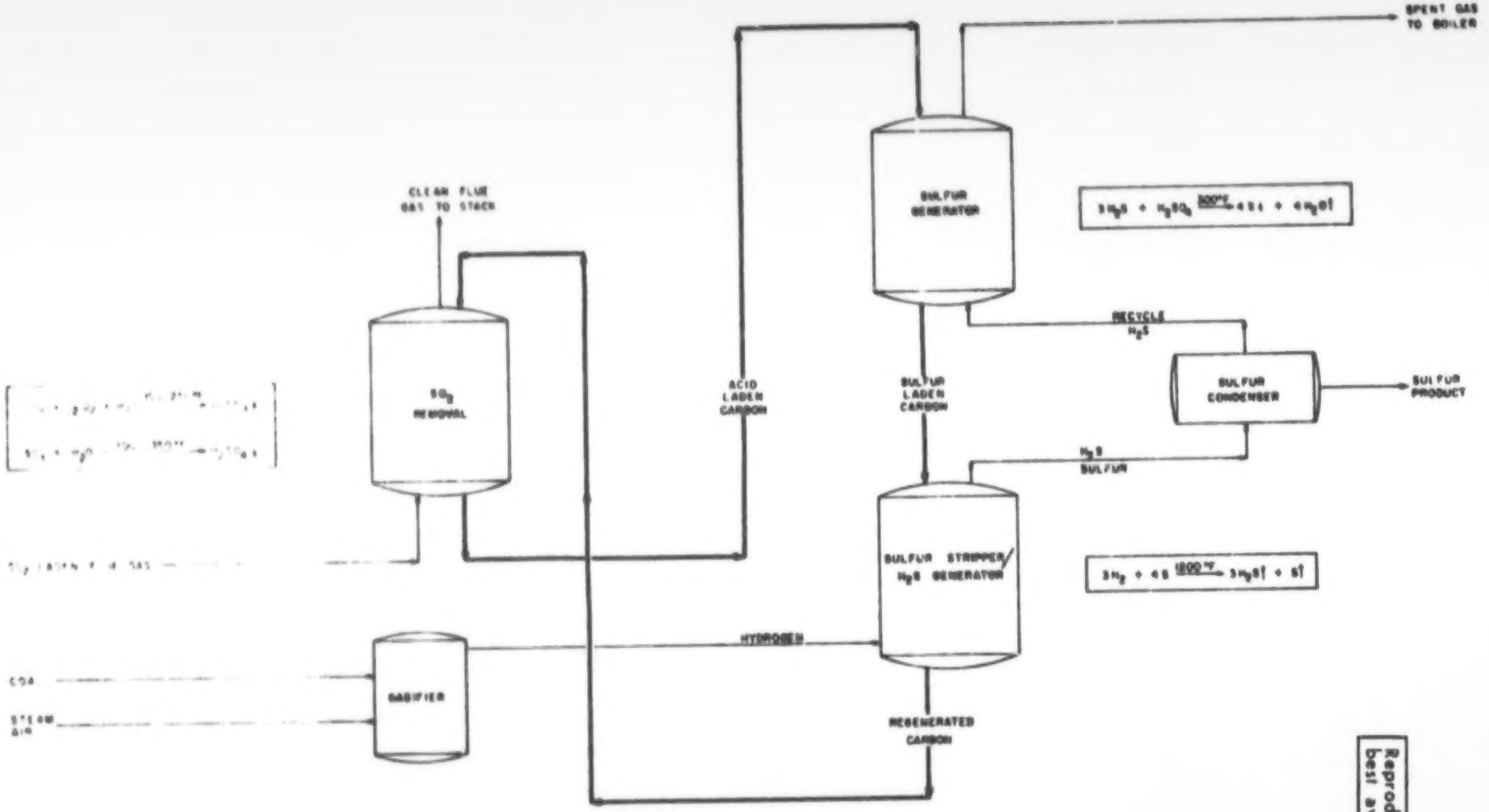

SPENT GAS TO BOILER
SULFUR GENERATOR
$3H_2S + H_2SO_4 \xrightarrow{300°F} 4S + 4H_2O\uparrow$
RECYCLE H_2S
SULFUR CONDENSER
SULFUR PRODUCT
H_2S SULFUR
SULFUR LADEN CARBON
SULFUR STRIPPER/ H_2S GENERATOR
$3H_2 + 4S \xrightarrow{1200°F} 3H_2S\uparrow + S\uparrow$
CLEAN FLUE GAS TO STACK
SO_2 REMOVAL
ACID LADEN CARBON
SO_2 LADEN FLUE GAS
HYDROGEN
GASIFIER
COAL
STEAM AIR
REGENERATED CARBON

FIGURE B-11. WESTVACO SO_2 REMOVAL PROCESS(92)

Reproduced from best available copy.

condenser where liquid sulfur at 270°F is separated from the H_2S. The sulfur is filtered to remove dust, solidified, and stored. H_2S is recycled to the sulfur generator. The regenerated carbon is cooled to 300°F and returned to the absorber.

Removal Efficiency

The system is capable of removing 90 percent SO_2 from flue gas.

By-Product

Elemental sulfur (99.7 percent pure) is the by-product. A 15-MW boiler, generating about 30,000 scfm of 3,300 ppm SO_2 flue gas would produce about 419 lbs/hr of sulfur.(93)

Wastes

None.

Heat and Material Requirements

Some carbon is lost due to mechanical attrition, normally less than 1 percent per cycle. A 15-MW installation, circulating 6,600 lbs carbon/hr would require about 25 lbs carbon/hr of makeup. Coal would be consumed in the gasifier to generate the H_2 reduction stream. About 716 lbs of coal/hr would be required for the 15-MW system. Sixty three gallons/hr of No. 2 fuel oil would be consumed in the H_2S generator/sulfur stripper.

Advantages

(1) Production of marketable by-product, elemental sulfur
(2) High SO_2 removal efficiency (90 percent)
(3) No stack gas reheat
(4) No waste stream generated.

Disadvantages

(1) Extensive solids handling

(2) Complicated process

(3) Requirements of hydrogen and fuel oil in the process.

Development Status

The Westvaco process was originally designed for Claus plant tail gas applications. Westvaco has recently completed pilot plant tests on a 20,000 ft^3/hr flue gas stream from an oil-fired boiler. They are interested in evaluating their system on a coal-fired boiler and have designed a 15-MW prototype unit. They are actively pursuing paths for the installation of such a unit.

Capital and Operating Costs

The capital cost for a 15-MW battery limit installation was estimated at $\$2.4 \times 10^4$ by Westvaco in August, 1974.(93) The cost included construction expense, contractor's fee, and contingency. The labor, material, and utility requirements for the installation are shown in Table B-15.(93)

TABLE B-15. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR WESTVACO PROCESS(93)

Basis: 15-MW coal-fired boiler
Flue gas flow rate: 30,000 acfm
SO_2 concentration, 3,230 ppm

Item	Quantity
Utility	
Power	670 kW
Fuel oil (No. 2)	63 gal/hr
Steam	12,000 lbs/hr
Cooling water	500 gpm
Material	
Activated carbon	25 lbs/hr
Coal	0.358 tons/hr
Labor	
Direct operation	2 men/shift(a)
Maintenance	4 percent of capital cost(b)

(a) Engineer, 1 man/shift; technician, 1 man/shift.

(b) This value was assumed.

Sulfacid Process

Developer/Manufacturer

The Sulfacid process was developed by Lurgi of Frankfurt, West Germany, and is licensed in the United States to the Rust Engineering Company, Birmingham, Alabama.

Process Description

The Sulfacid process utilizes an impregnated carbon bed to adsorb and oxidize sulfur dioxide, and water to reactivate the bed.(94) As can be seen in Figure B-12, stack gas is pretreated to adjust the temperature, humidity, and particle content in a humidifying chamber or a venturi scrubber. The conditioned gas with a temperature of 120°F to 175°F, a dew point of about 120°F, and particle loading of less than 0.007 grain/scf flows upward at low velocity through a bed of carbon-based catalyst of 1 to 2 feet deep. Sulfur dioxide, oxygen, and water are adsorbed on the impregnated carbon where sulfuric acid is formed by the reaction

$$SO_2 + 1/2\ O_2 + H_2O \rightarrow H_2SO_4.$$

The acid is washed from the bed by a continuous spray of water as it is formed. The product acid flows continuously from the reactor as a 10 to 15 percent solution. This solution can be used to quench high temperature gas streams, thereby increasing its concentration up to 20 to 30 percent, if desired.

Removal Efficiency

A sulfur dioxide removal efficiency of 90 percent can be achieved by the process. Sulfur dioxide removal efficiency is a function of catalyst depth, making the system amenable to efficiency upgrading, if necessary, after installation.

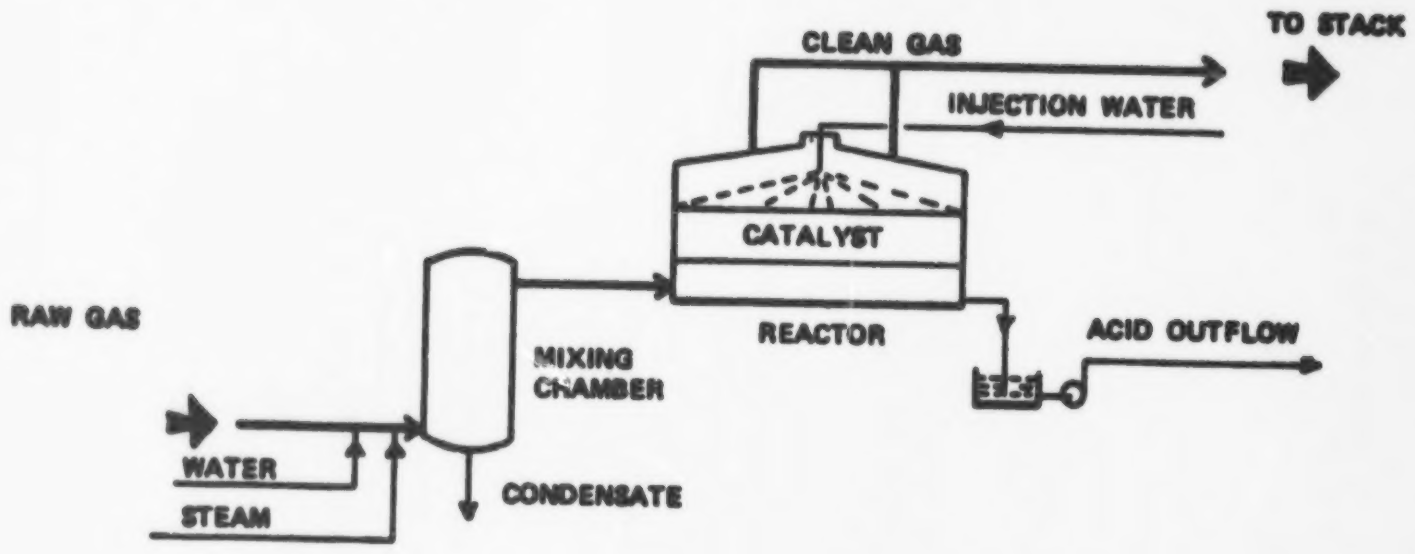

CLEAN GAS
TO STACK
INJECTION WATER
CATALYST
RAW GAS
REACTOR
ACID OUTFLOW
MIXING CHAMBER
WATER
STEAM
CONDENSATE

FIGURE B-12. SULFACID PROCESS(94)

Material and Heat Requirements

For the flue gas characterized in Table B-16, about 55.2 gal/min and 42.5 gal/min of process water are required to generate 12.5 percent and 25 percent acids by weight, respectively. The material balance was based on a 90 percent sulfur dioxide removal efficiency. Heat is required to reheat the flue gas. About 1,100 Btu are consumed per 1000 scfm of flue gas flow rate.

By-Product

The Sulfacid system handling the flue gas characterized in Table III-13 produces about 152.4 ton/day of 12.5 percent sulfuric acid or 76.2 ton/day of 25 percent sulfuric acid. If there is no market available for the acid, it should be neutralized with limestone for disposal.

Wastes

Residual sulfur dioxide emission is the only emission resulting from the process. If the by-product sulfuric acid should be neutralized by limestone or lime for disposal, the resulting waste sludge for the flue gas in Table B-16 would be about 52.9 ton/day (solids content of 50 percent).

Advantages

The following advantages of the process have been reported.(94,95)

(1) Simple regeneration
(2) Low operating cost (simple operation)
(3) High reliability (less moving parts)
(4) Adjustable removal efficiency
(5) Simple in overall process.

TABLE B-16. FLUE GAS CHARACTERISTICS FOR SULFACID PROCESS

Item	Value
Flow rate	37,900 scf/min
Temperature	510°F
SO_2 concentration	3,000 ppm
Water vapor content	7.3% by volume

Disadvantages

The following disadvantages of the process have been reported.(94,95)

(1) Generation of low concentration by-product acid to be used or disposed of

(2) High water consumption

(3) Possible corrosion

(4) Potential cold, wet plume problem.

Development Status

The Sulfacid process has been commercially applied to the treatment of chemical plant waste streams for several years in Europe. One system in West Germany has been installed on a titanium dioxide recovery process with a flue gas flow rate of about 20,000 cfm. Another system in Holland has been installed on a sulfuric acid plant with a flue gas rate of 20,000 cfm. No major problems have been encountered regarding catalytic activity of impregnated carbon, and general operation and maintenance for 7 years. A plant to handle 20,000 to 30,000 cfm of sulfuric acid plant tail gas will be built in Pittsburgh, Pennsylvania for the United States Steel Company. In general, the Sulfacid process lends itself more to sulfuric acid plant tail gases because of the high volume of low concentration sulfuric acid by-product.

Capital and Operating Costs

The total capital cost for a Sulfacid system handling a flue gas of 72,000 acfm was quoted as about $2 million.(95) This excluded the cost for a reheat system and assumed a low fly ash concentration in the flue gas. The adjusted cost, including the cost for a reheater was given as $2.1 million. The corresponding cost for a Sulfacid system handling a flue gas of 7,200 acfm was estimated at $520,000.(95) The exponential scale factor for total capital cost was estimated to be 0.6 and the construction time required to be about 6 months for custom design.(95) The labor, material, and utility requirements of the system handling of the flue gas listed in Table B-16 are shown in Table B-17.

TABLE B-17. LABOR, MATERIALS, AND UTILITY REQUIREMENTS FOR SULFACID PROCESS

Item	Quantity
Utilities	
Power	170 kW(a)
Process water	55.2 gpm for 12.5 percent by-product acid 42.5 gpm for 25 percent by-product acid
Steam (for reheat)	9,500 lb/hr(b)
Material	
Limestone(c)	25.2 tons/day(d)
Makeup carbon	nil
Labor	
Direct operating	1 man/shift(e)
Maintenance	1 percent of capital cost(e)

(a) The value was derived from Reference 94.

(b) This was obtained from Reference 94.

(c) It was assumed that limestone is used to neutralize the by-product acid.

(d) This was obtained based on the use of 10 percent excess limestone with a purity of 85 percent.

(e) The value was obtained from the Rust Engineering Company.(95)

Chiyoda Process

Developer/Manufacturer

The Chiyoda Chemical Engineering and Construction Company, Japan, developed the "Thoroughbred 101" sulfur dioxide removal process. They have designed and manufactured the systems for Claus plants and oil-fired boilers in Japan, and the process is marketed in the United States through their Seattle, Washington office.

Process Description

Following treatment by an ESP for fly ash removal, the flue gas is introduced into a venturi-type prescrubber where the gas is cooled and any remaining particulates are removed (see Figure B-13). The gas passes through a packed bed absorber where it flows upward contacting a counter-current 2 to 5 percent sulfuric acid solution; containing about 2,000 ppm of ferric sulfate. The SO_2 is absorbed by the acid solution

$$SO_2 + H_2O \rightarrow H_2SO_3 \quad .$$

Pressure drop through the venturi and scrubber is about 9 inches water and the liquid-to-gas ratio in the scrubber is about 300 gal/1000 scf. After passing through a demister, the flue gas is reheated and vented to the stack.

The scrubber effluent solution flows to the oxidizer tower, where, in the presence of the ferric ion catalyst, air injected into the liquor oxidizes the H_2SO_3 to H_2SO_4.

$$H_2SO_3 + 1/2\ O_2 \rightarrow H_2SO_4$$

A portion of the liquor leaving the oxidizer returns to the absorber, and the remainder passes to gypsum production. First, the liquor is neutralized with lime or pulverized limestone to form insoluble gypsum.

$$H_2SO_4 + CaCO_3 + H_2O \rightarrow CaSO_4 \cdot 2H_2O + CO_2$$

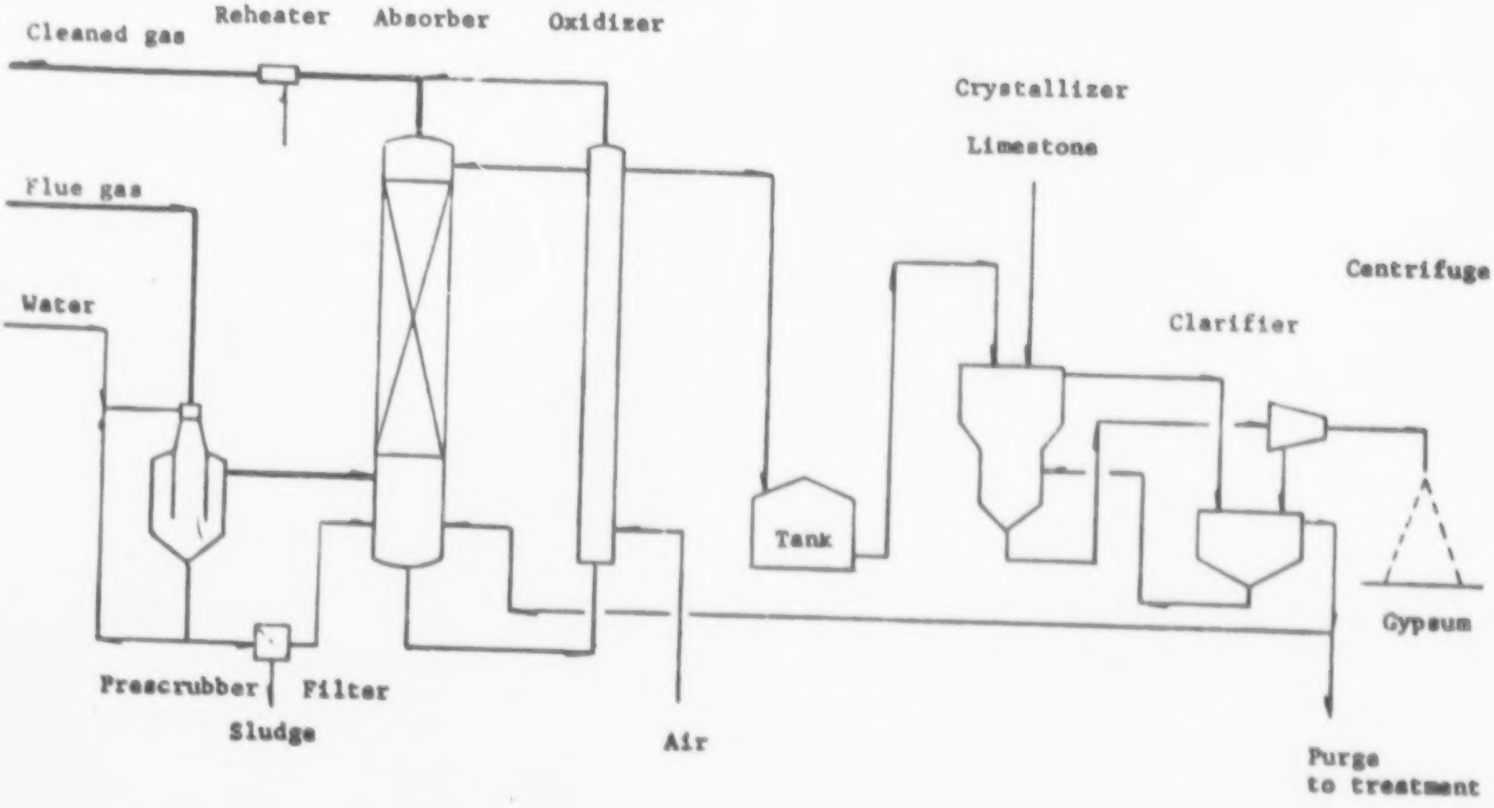

Cleaned gas
Reheater
Absorber
Oxidizer
Flue gas
Water
Crystallizer
Limestone
Centrifuge
Clarifier
Tank
Gypsum
Prescrubber
Filter
Sludge
Air
Purge
to treatment

FIGURE B-13. PROCESS FLOW DIAGRAM OF CHIYODA PROCESS(96)

Then, gypsum crystals are withdrawn from the bottom of the crystallizer and centrifuged. Crystallizer overflow and mother liquor from the centrifuge are sent to a clarifier. The clarifier underflow is returned to the crystallizer and the overflow is recycled to the absorber. The centrifuged gypsum contains about 5 to 20 percent free water. It may be trucked away and used in wallboard production.

Chloride ion accumulation in the scrubber corrodes the stainless steel and a purge stream must be withdrawn to maintain the chloride content below 200 ppm.

Removal Efficiency

The process is capable of a 95 percent SO_2 removal efficiency.

Raw Material and Heat Requirements

Heat is required to reheat the flue gas. Ferric sulfate is lost with the gypsum and purge stream and must be replaced. Limestone is consumed in the crystallizer.

By-Product

About 2.44 lbs of gypsum containing 85 to 90 percent solids ($CaSO_4 \cdot 2H_2O$, 97 percent; limestone, 0.6 percent; others, 2.4 percent) is produced per pound of SO_2 removed.

Wastes

The fly ash filter cake and purge stream are emitted from the process.

Advantages

(1) Relatively simple process

(2) Proven performance on oil-fired boilers, large and small
(3) Production of gypsum instead of sludge
(4) Good SO_2 removal efficiency
(5) No plugging or scaling in scrubber.

Disadvantages

(1) Flue gas reheat
(2) Poor market for gypsum in the U.S.
(3) Corrosion problem due to chloride ion accumulation
(4) High liquid to gas ratio and high pumping requirement
(5) Large absorber required.

Development Status

Since 1972, 10 commercial Chiyoda Thoroughbred processes have been installed on Claus plants and oil-fired boilers in Japan. A 23-MW pilot plant on Gulf Power Company's coal-fired Scholz plant in Sneads, Florida will be completed in late 1974.

Capital and Operating Costs

The capital cost for a 250-MW Chiyoda process was estimated at \$80 to \$100/kW.(96) For a Chiyoda process installed on a 30-MW boiler burning high sulfur resid (3.5 percent sulfur), the labor, material, and utility requirements are shown in Table B-18.(97)

TABLE B-18. LABOR, MATERIAL, AND UTILITY REQUIREMENTS AND BY-PRODUCT PRODUCTION FOR CHIYODA PROCESS(97)

Basis: Oil-fired boiler (30 MW equivalent)
Sulfur content of oil, 3.5 percent
Flue gas, 53,000 scfm
SO_2 concentration, 1,450 ppm
Gas temperature, 340 F

Item	Quantity
Utility	
Power	760 kW
Steam	11,900 lb/hr
Water	25 gpm[(a)]
Material	
Limestone	0.53 tons/hr
Labor	
Direct operation	1 man/shift
Maintenance	3.5 percent of capital cost[(a)]
By-Product	
Gypsum (10 percent free water)	0.95 tons/hr

(a) This value was assumed.

Ammonia Scrubbing Process (Peabody)

Developer/Manufacturer

The Peabody Engineering Company, Stanford, Connecticut, designs and constructs ammonia-based sulfur dioxide removal systems.

Process Description

Small industrial boiler facilities are not equipped to process the by-products, sulfur dioxide, and ammonium sulfite generated by thermal stripping or acidification of the spent ammonium scrubbing slurry. The following process is a proposed method for ammonium scrubbing with the generation of a marketable, easily handled by-product ammonium sulfate fertilizer.(98) The complete integrated process has not been tested, but the individual process operations (absorption, oxidation, and evaporation) are straightforward and are in common use in the ammonia industry.

The flue gas is first treated with an electrostatic precipitator or mechanical separator to remove the fly ash and particulate matter. It is cooled from 300 to 170°F in a water-cooled heat exchanger and introduced into the bottom of a four-stage absorber (see Figure B-14). First the remaining fly ash is scrubbed with water and the solids slurry is withdrawn and transported to a settling pond. The gas flows through the next three trays where it is scrubbed countercurrently with a mixture of ammonia, ammonium sulfate, ammonium bisulfite, and ammonium sulfate. The sulfur dioxide in the gas reacts with ammonia and ammonium sulfite in the solution to form ammonium bisulfite; oxidation of ammonium sulfite to ammonium sulfate also occurs.

$$NH_3 + H_2O + SO_2 = NH_4HSO_3$$

$$2NH_3 + H_2O + SO_2 = (NH_4)_2SO_3$$

$$(NH_4)_2SO_3 + SO_2 + H_2O \rightarrow 2NH_4HSO_3$$

$$(NH_4)_2SO_3 + \frac{1}{2}O_2 \rightarrow (NH_4)_2SO_4$$

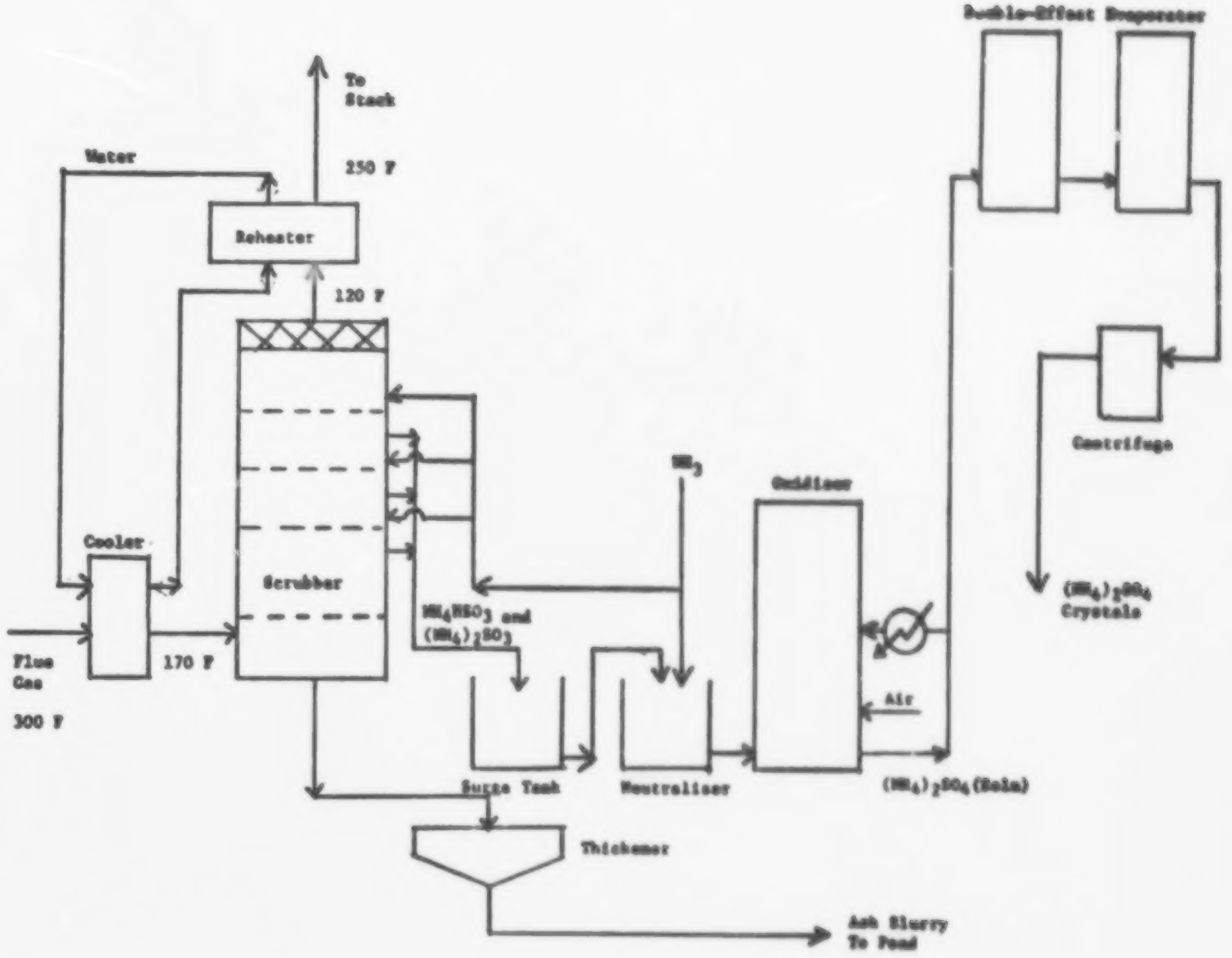

To Stack
250 F
Water
Reheater
120 F
Cooler
Scrubber
Flue Gas
300 F
170 F
NH_4HSO_3 and $(NH_4)_2SO_3$
NH_3
Surge Tank
Neutralizer
Thickener
Ash Slurry To Pond
Oxidizer
Air
$(NH_4)_2SO_4$ (Soln)
Double-Effect Evaporator
Centrifuge
$(NH_4)_2SO_4$ Crystals

FIGURE B-14. AMMONIA SCRUBBING WITH OXIDATION TO AMMONIUM SULFATE

Peabody impingement-type trays are used, with each tray having a separate circulation system. The flue gas passes upward through a mist eliminator to capture any vaporized ammonia, and exits through the top of the scrubber. It is reheated from 120 to 250°F by passing through a water-heated heat exchanger; cooled water from the heat exchanger is recycled to the flue gas precooler heat exchanger. The heated effluent gas is vented to the stack.

Spent absorbent slurry containing about 50 percent ammonium salts is withdrawn from each tray and pumped to a surge tank. From the surge tank the slurry is pumped to the neutralizer where ammonia is added to convert the bisulfite to sulfite to minimize sulfur dioxide loss during oxidation.

$$NH_3 + NH_4HSO_3 \rightarrow (NH_4)_2SO_3$$

Water is also added to prevent ammonium sulfate crystallization in the oxidizer. The neutralized solution is introduced to the oxidizer where the sulfite is reacted with air at 100 psig and 185°F to produce the sulfate.

$$(NH_4)_2SO_3 + 1/2\ O_2 \rightarrow (NH_4)_2SO_4$$

Heat of reaction is removed by circulating the solution through a water-cooled heat exchanger to maintain the temperature at 185°F. Temperature control is necessary to maintain the solubility of oxygen. A slip stream, with about 40 percent solids, is withdrawn from the cooling stream and pumped to a double-effect vacuum evaporator crystallizer. The crystals are separated in a centrifuge and the liquid phase returned to the evaporator. The ammonium sulfate cake (70 percent solids) may be conveyed to a storage bin or dried in a spray dryer.

Removal Efficiency

Greater than a 90 percent removal efficiency of sulfur dioxide can be obtained with ammonia scrubbing. The mechanical separator or low-efficiency ESP coupled with the scrubber can remove up to 99.5 percent of fly ash and particulate matter. The system has no NO_x removal capability.

Raw Material and Heat Requirements

Ammonia and process steam are required. Utilizing a water recirculated gas cooling and heating system, the heat requirements for reheating the flue gas can be eliminated. Steam, however, will be required in the double-effect evaporator.

By-Product

The usable by-product from the process is ammonium sulfate.

Wastes

The only solid waste is the fly ash removed from the scrubber. If the scrubber system is not equipped with a highly effective demister, the stack gas would contain considerable amounts of ammonium sulfite compound.

Advantages

(1) Good SO_2 removal efficiency, 90 percent or greater

(2) No scaling or plugging problems in scrubber

(3) Production of marketable by-product, ammonium sulfate fertilizer

(4) No regeneration of absorbent required.

Disadvantages

(1) High cost of ammonia absorbent and questionable availability

(2) Unstable market for ammonium sulfate fertilizer

(3) Flue gas reheat system required

(4) Loss of ammonia to stack and resultant "blue" plume requires investment in costly Brinks mist eliminator.

Development Status

The process has not been tested for flue gases from coal-fired boilers. However, the individual process operations have been utilized in the ammonia industry for several years. The absorption unit is manufactured and marketed by Peabody Engineering and is used in several of their ammonium sulfur dioxide removal systems. The Japan Engineering Consulting Company (JECCO) developed the oxidation system; it has been proven in large installations in Japan. Peabody Engineering designed and constructed ammonium scrubbing systems primarily for paper mills where the regenerated ammonium sulfite is used in the cooking liquor (see Table B-19).

Capital and Operating Costs

The capital cost information was obtained from the result of the TVA study.(99) The estimated cost for a 200-MW coal-fired boiler system was \$5.089 x 10^6 in 1969. The labor, material, and utility requirements are shown in Table B-20.

TABLE B-19. PEABODY AMMONIA SCRUBBING IN COMMERCIAL OPERATION

Plant/ Location	Type of Gas	Flow Rate, acfm	Gas Temperature, F	SO_2 Inlet ppm	SO_2 Outlet ppm	Remarks
Conserv. Chemicals, Barton, Florida	Sulfuric acid Tail gas	108,000	195	2,025	250	Scrubber effluent is used in fertilizer production. Startup in spring, 1974.
Boise Cascade, Salem, Oregon	Flue gas from black liquor	122,000	450	8,000	800	System produces $(NH_4)_2SO_3$ pulp cooking liquor. Startup in 1972.
Rayonier Quebec, Port-Cartier, Quebec	Flue gas from black liquor	400,000	470	--	200	System produces $(NH_4)_2SO_3$ pulp cooking liquor. Startup expected soon.
USF Agri-Chemicals, Ft. Meade, Florida	Sulfuric acid Tail gas	174,000	220	--	--	Contract signed.

TABLE B-20. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR PEABODY AMMONIA SCRUBBING PROCESS(99)

Basis: 200-MW boiler
3.5 percent sulfur in coal
554,400 tons/yr coal burned
7,000 hrs/yr operation

Item	Quantity
Utility	
Power	3.27 MW
Steam	10,900 lb/hr
Water	1,848 gpm
Material	
Ammonia	1.58 tons/hr
Labor	
Direct operation	1.45 man/shift
Maintenance	20 percent of capital cost/yr

Ammonia Scrubbing Process (Catalytic, Inc.)

Developer/Manufacturer

Catalytic, Inc., a subsidiary of Air Products and Chemicals, Inc., developed the ammonia-based SO_2 scrubbing system. The system utilizes the Institut Francais du Petrole (IFP) reducing process to regenerate the spent ammonium salts and produce sulfur. The integrated process is marketed in the United States by Catalytic.

Process Description

Following treatment by an ESP for fly ash removal, the flue gas is cooled and water saturated by passing through a venturi scrubber (see Figure B-15). The water-fly ash slurry is thickened, neutralized with lime, and either filtered or pumped to a disposal facility. From the venturi the flue gas flows upward through the absorber where it contacts a 14 mole percent slurry of $(NH_4)_2SO_3$, $(NH_4)_2SO_4$, and NH_4HSO_3. The following reactions occur:

$$SO_2 + (NH_4)_2SO_3 + H_2O = 2NH_4HSO_3$$

$$SO_3 + (NH_4)_2SO_3 = (NH_4)_2SO_4 + SO_2$$

$$O_2 + 2(NH_4)_2SO_3 = 2(NH_4)_2SO_4.$$

There is normally about 10 percent oxidation of ammonium sulfite to ammonium sulfate.

The adsorber is a cylindrical column containing 3 or 4 floating cap trays, each with an individual circulation system. Ammonia concentration, pH, and liquor to gas ratio (normally about 5 gal/1000 cf) in the column are controlled to eliminate the "blue" plume from the scrubbing systems. The purified gas flows from a mist eliminator through a blower and to the stack. About 5 ppm of ammonium sulfite compounds exit with the flue gas

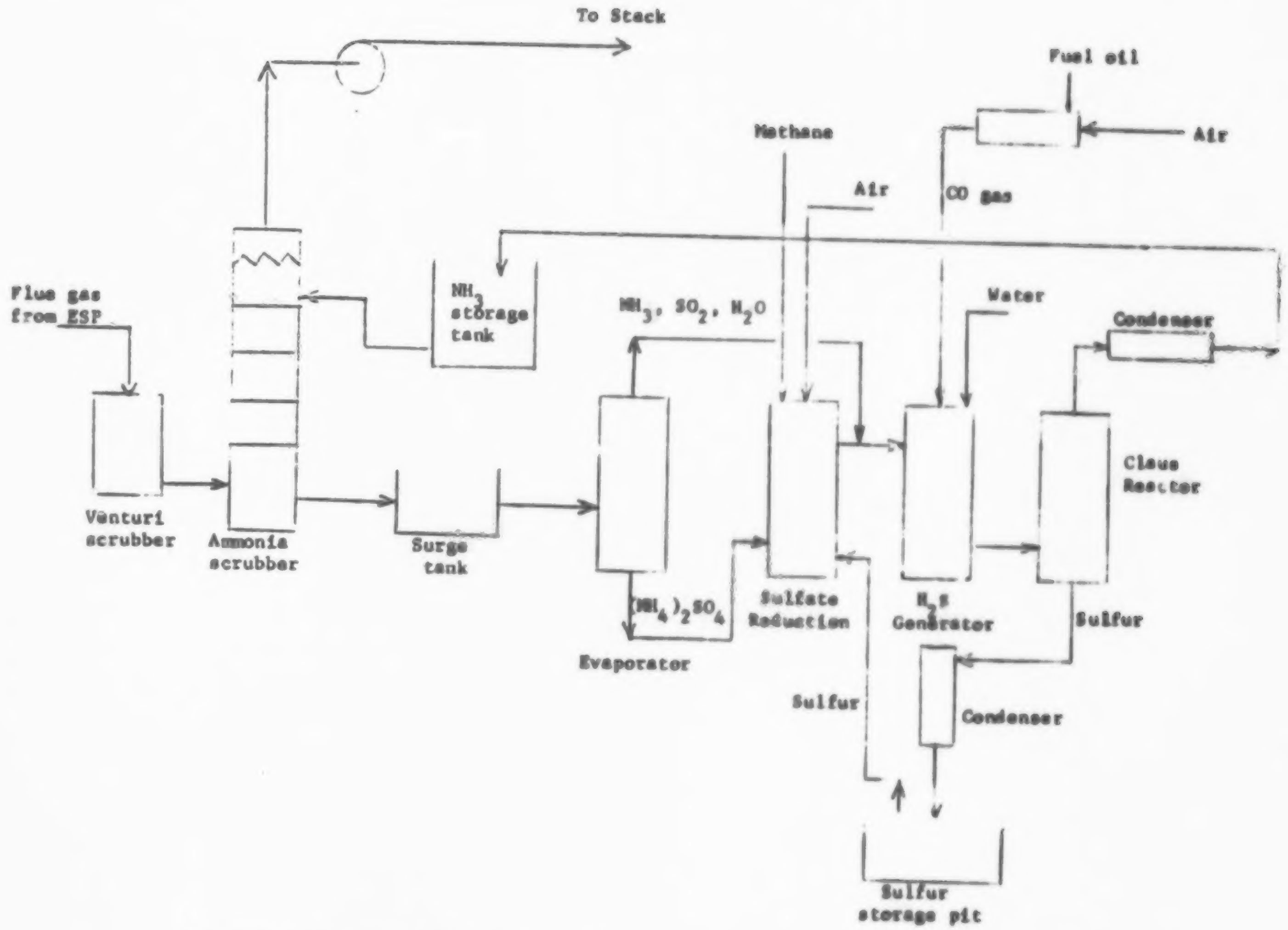

To Stack
Fuel oil
Air
Methane
Air
CO gas
Flue gas from ESP
NH_3 storage tank
NH_3, SO_2, H_2O
Water
Condenser
Venturi scrubber
Ammonia scrubber
Surge tank
Evaporator
$(NH_4)_2SO_4$
Sulfate Reduction
H_2S Generator
Claus Reactor
Sulfur
Condenser
Sulfur
Sulfur storage pit

FIGURE B-15. FLOW PROCESS DIAGRAM - CATALYTIC AMMONIA SCRUBBING PROCESS(100)

The spent scrubbing slurry is pumped from the bottom of the absorber to a surge tank. From the surge tank it is pumped to an evaporator where at 300 F and 35 psi the less stable sulfite and bisulfite are converted to SO_2 and NH_3.

$$(NH_4)_2SO_3 = SO_2 + H_2O + 2NH_3$$

$$NH_4HSO_3 = SO_2 + H_2O + NH_3$$

The SO_2 and NH_3 gases are transferred to the H_2S generating vessel and the ammonium sulfate crystals are discharged to the sulfate reduction reactor. Methane is burned to provide process heat maintaining the reduction reactor at 700 F. Sulfur is added to produce the SO_2 and NH_3.

$$(NH_4)_2SO_4 = NH_4HSO_4 + NH_3$$

$$2NH_4HSO_4 + S = 3SO_2 + 2H_2O + 2NH_3$$

Carbon monoxide, generated by burning fuel oil, water, and hydrogen, are combined to produce the desired amount of H_2S for the Claus reaction

$$SO_2 + 3CO + H_2O = H_2S + 3CO_2$$

$$SO_2 + 2H_2 = H_2S + 2H_2O.$$

In the Claus reactor the SO_2 and H_2S are combined in a 1 to 2 ratio to generate elemental sulfur.

$$2H_2S + SO_2 = 3S + 2H_2O$$

The sulfur is condensed and pumped to a storage pit where a portion of it is withdrawn for use in the sulfate reduction reaction. The ammonia and water off gases are condensed to a liquid and recycled to the scrubber.

Removal Efficiency

The process is capable of removing 90 percent of the SO_2 from flue gas.

By-Product

Elemental sulfur is produced from the process.

Wastes

A fly ash slurry waste stream is generated in the venturi prescrubber.

Raw Material and Heat Requirements

Ammonia makeup is required to replace losses to the flue gas and in the regeneration system. Fuel oil is necessary to generate the reducing gas in the manufacture of H_2S. Steam heat is required in the evaporator and sulfur generating systems.

Advantages

(1) Good SO_2 removal efficiency
(2) No waste sludge disposal problem
(3) Generation of marketable by-product sulfur
(4) No plugging or scaling.

Disadvantages

(1) High cost of ammonia
(2) High capital cost
(3) Entire system not proven on utility boiler.

Development Status

In 1970 IFP began marketing its Claus tail gas treating process. It involves ammonia scrubbing coupled with the reduction-regeneration system. To date, seven installations have been constructed and all are currently operating. The complete IFP SO_2 removal system is being installed in a 35-MW utility boiler in France.

In 1972 Catalytic, Inc., TVA, and EPA jointly evaluated the feasibility of Catalytic, Inc.'s ammonia scrubbing system at TVA's Colbert Station pilot plant. Catalytic guarantees the continuous supply of ammonia to their customers and is actively engaged in marketing their ammonia-based scrubbing system coupled with the IFP regeneration process.

Capital and Operating Costs

The capital cost for a 20-MW coal-fired boiler (3.5 percent sulfur in coal) producing a flue gas of 42,000 scfm (2,500 ppm SO_2) was estimated at $3 million in November, 1974.(101) The cost was for the complete battery limit system including engineering, royalties, site development, construction, etc.

The annual operating cost for the 20-MW system was estimated at $550,000 in November, 1974, not including the credit for by-product sulfur. The detailed breakdown of labor, material, and utility requirements are not available.

Shell Flue Gas Desulfurization Process (Shell FGD)

Developer/Manufacturer

Shell International Petroleum developed the process in the early 1960's. Universal Oil Products (UOP) purchased the licensing rights for the U.S. from Shell in 1971 and is marketing the process to chemical plants and utility boilers.

Process Description

The Shell FGD process employs a copper oxide-alumina adsorbent in a fixed-bed reactor to adsorb SO_2 from the flue gas. The spent adsorbent is regenerated with a stream of 30 percent H_2 in steam to generate an off-gas stream which can be processed to yield liquid SO_2, sulfuric acid, or elemental sulfur.

After passing through the economizer and a particulate removal system, the flue gas is passed through a blower and into the adsorber (see Figure B-16). The adsorber is a fixed bed in which the flue gas flows through open channels along side and in contact with the adsorbent material. The absorbent is elemental copper supported on an alumina structure and contained in unit cells.

Upon contact with the flue gas, the copper and any cuprous sulfide (Cu_2S) contaminants are oxidized to CuO and $CuSO_4$.

$$Cu + 1/2\ O_2 \rightarrow CuO$$

$$Cu_2S + 5/2\ O_2 \rightarrow CuO + CuSO_4$$

The existence of Cu_2S is undesirable as one-half of the copper is converted to $CuSO_4$ and is unavailable to participate in the acceptance reaction. The SO_2 reacts with the CuO to form $CuSO_4$.

$$CuO + 1/2\ O_2 + SO_2 \rightarrow CuSO_4$$

When the adsorber becomes loaded with sulfur, it is arranged for regeneration. A stream of hydrogen is passed through the bed and the conversion of $CuSO_4$ to Cu, H_2O, and SO_2 takes place.

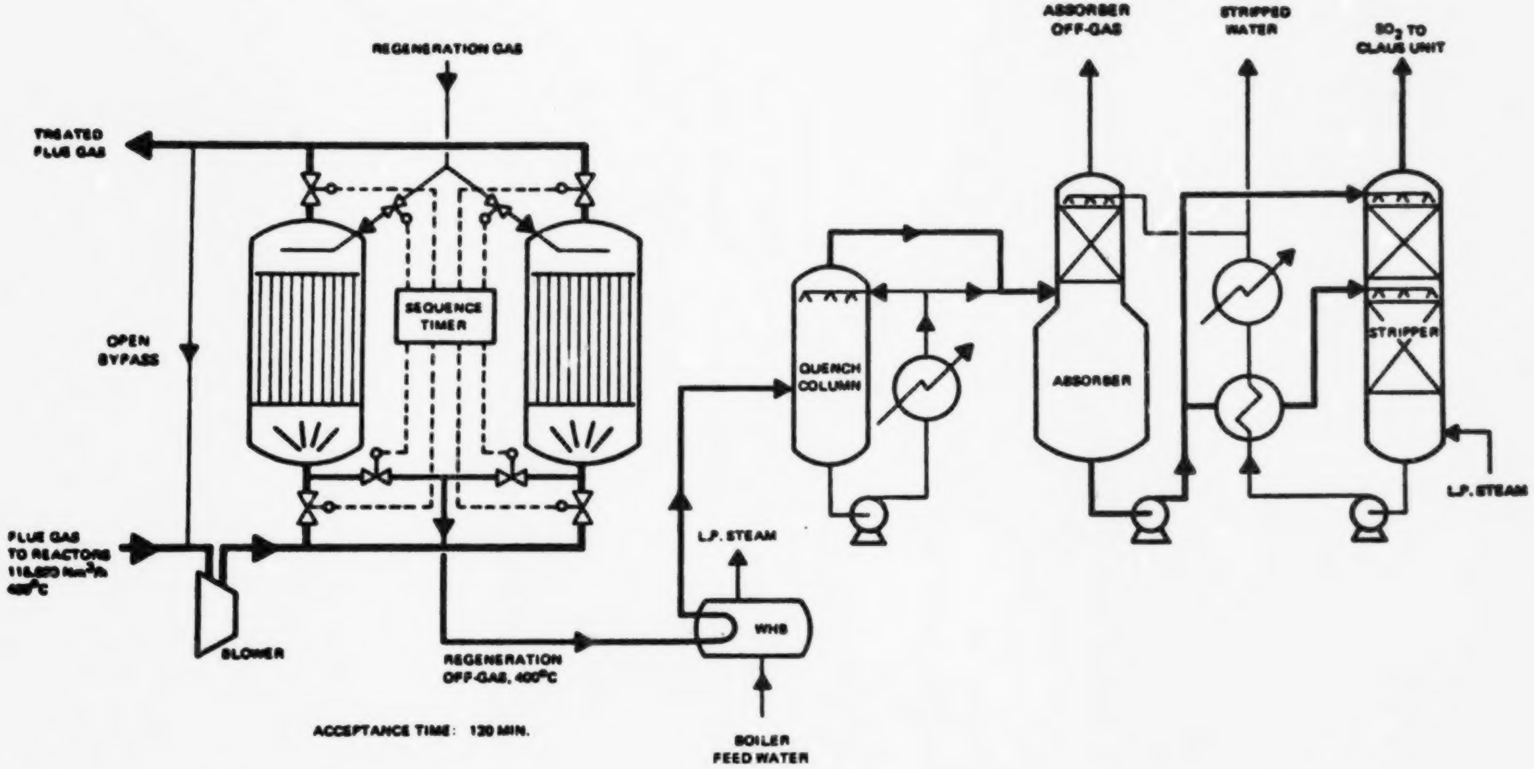

REGENERATION GAS
TREATED FLUE GAS
OPEN BYPASS
SEQUENCE TIMER
FLUE GAS TO REACTORS
BLOWER
REGENERATION OFF-GAS, 400°C
ACCEPTANCE TIME: 120 MIN.
L.P. STEAM
WHB
BOILER FEED WATER
QUENCH COLUMN
ABSORBER OFF-GAS
ABSORBER
EXCESS STRIPPED WATER
SO_2 TO CLAUS UNIT
STRIPPER
L.P. STEAM

FIGURE B-16. SIMPLIFIED PROCESS FLOW SCHEME OF SHELL FGD UNIT(102)

$$CuO + H_2 \rightarrow Cu + H_2O$$

$$CuSO_4 + 2H_2 \rightarrow Cu + SO_2 + 2H_2O$$

Two or more identical adsorbers are applied in cyclic operation to provide for continuous processing of flue gas. The off gas SO_2 stream is further concentrated to 90 percent by removing the H_2O and inerts and may be either liquefied or processed in a Claus reactor to produce elemental sulfur.

Removal Efficiencies

The Shell FGD process is capable of removing 90 percent of the SO_2 from flue gas. The process may also remove NO_x. It has no particulate removal capability.

Raw Material and Heat Requirements

About 0.1 lb of H_2 is required in the regeneration process for every pound of SO_2 recovered.(102) Heat is required to raise the temperature of the flue gas to the optimum reaction temperature of 700 F. With the recovery of some portion of the added thermal energy, the net reheat requirement is approximately 1 percent of the fuel input to the boiler.

By-Product

The by-product is elemental sulfur.

Waste

The only waste is stripping water containing 20 ppm S, of which 75 percent is present as sulfate.

Advantages

(1) Dry process, no flue gas reheat

(2) No waste sludge generation

(3) Reliable operation
(4) Low utility requirement.

Disadvantages

(1) Requires SO_2 processing plant
(2) Expensive for small systems
(3) Requires H_2 reduction gas
(4) Requires "hot" ESP
(5) Requires expensive reheat for retrofit.

Development Status

In the early 1960's Shell developed the Shell Flue Gas Desulfurization process. In 1967 a 400-600 scfm side stream was withdrawn from the flue gas of an oil-fired boiler at the Shell refinery near Rotterdam, Netherlands to evaluate reactor design, catalyst type, and operating parameters. The system operated for approximately 20,000 hours. In mid-1973, in Japan, a commercial-size system was constructed to process the combined flow of flue gas from an oil-fired boiler and a Claus plant tail gas stream, a total flow of about 90,000 scfm. It is reportedly operating satisfactorily.

Following short tests on a coal-fired boiler in Rotterdam to assess the effects of fly ash on acceptor life, a processing system was installed on a 1400 scfm slip stream at Tampa Electric Company's Unit No. 1 coal-fired boiler. The purpose of the tests are to evaluate acceptor life under adsorption-regeneration cycling. The test module consists of only one reactor; bottled hydrogen is used to regenerate the acceptor and the regenerated off gas is vented to the stack. Testing began summer of 1974 and is expected to be completed by spring, 1975.

Capital and Operating Costs

The capital cost of a Shell FGD system for a small coal-fired boiler (i.e., capacities less than 40 MW) was estimated at about $100/kW capacity in 1974.(103) This included costs for the elemental sulfur reduction system. The operating labor, material, and utility requirements for a Shell FGD system installed on an oil-fired boiler (30 MW) are shown in Table B-21.(104) This included costs for the elemental sulfur recovery process operation.

TABLE B-21. LABOR, MATERIAL, AND UTILITY REQUIREMENTS FOR SHELL FGD PROCESS(104)

Basis: 30-MW oil-fired boiler
Sulfur content, 2.85 percent
88 percent SO_2 removal efficiency
Flue gas flow rate, 59,000 scfm
Onstream time, 8,000 hrs/yr

Item	Quantity
Utility	
Power	110 kW
Material	
Catalyst	\$4/hr(a)
Hydrogen	200 lb/hr
Labor	
Direct operation	0.5 men/shift
Maintenance	5 percent of capital cost

(a) 1972 cost.

Citrate Process

Developer/Manufacturer

The U.S. Bureau of Mines developed the Citrate Process for removing SO_2 from industrial waste gas. The Morrison-Knudsen Company, Inc., Boise, Idaho, and Peabody Engineering Company, independently offers the process on a commercial basis.

Process Description

The Citrate Process involves absorption of SO_2 by a solution of sodium citrate, citric acid, and sodium thiosulfate followed by reacting the absorbed SO_2 with H_2S to precipitate elemental sulfur and regenerate the citrate solution.

Following treatment by an ESP or cyclone to remove fly ash, the flue gas is passed through a humidifier (see Figure B-17). The humidifier is a fiberglass-lined tower containing a section packed with 1-inch saddles and a stainless steel mist eliminator. Water is used to cool the gas to about 140°F and eliminate H_2SO_4 mist and any remaining fly ash. The cleaned and cooled gas stream passes upward through the packed absorber where it contacts counter-currently a citrate solution (pH, 4.5). The SO_2 is removed by the following reactions.

$$SO_2 + H_2O = HSO_3^- + H^+$$

$$H^+ + HCit^= = H_2Cit$$

As the absorption of SO_2 is pH-dependent, decreasing with decreasing pH, the citrate functions as a buffering agent. The spent citrate solution flows to the sulfur reactor where H_2S gas is bubbled through the solution and reacts with the SO_2 in the aqueous solution. Although the chemistry of the reaction is complex, the overall reaction is as follows.

$$SO_2 + 2H_2S \rightarrow 3S + 2H_2O$$

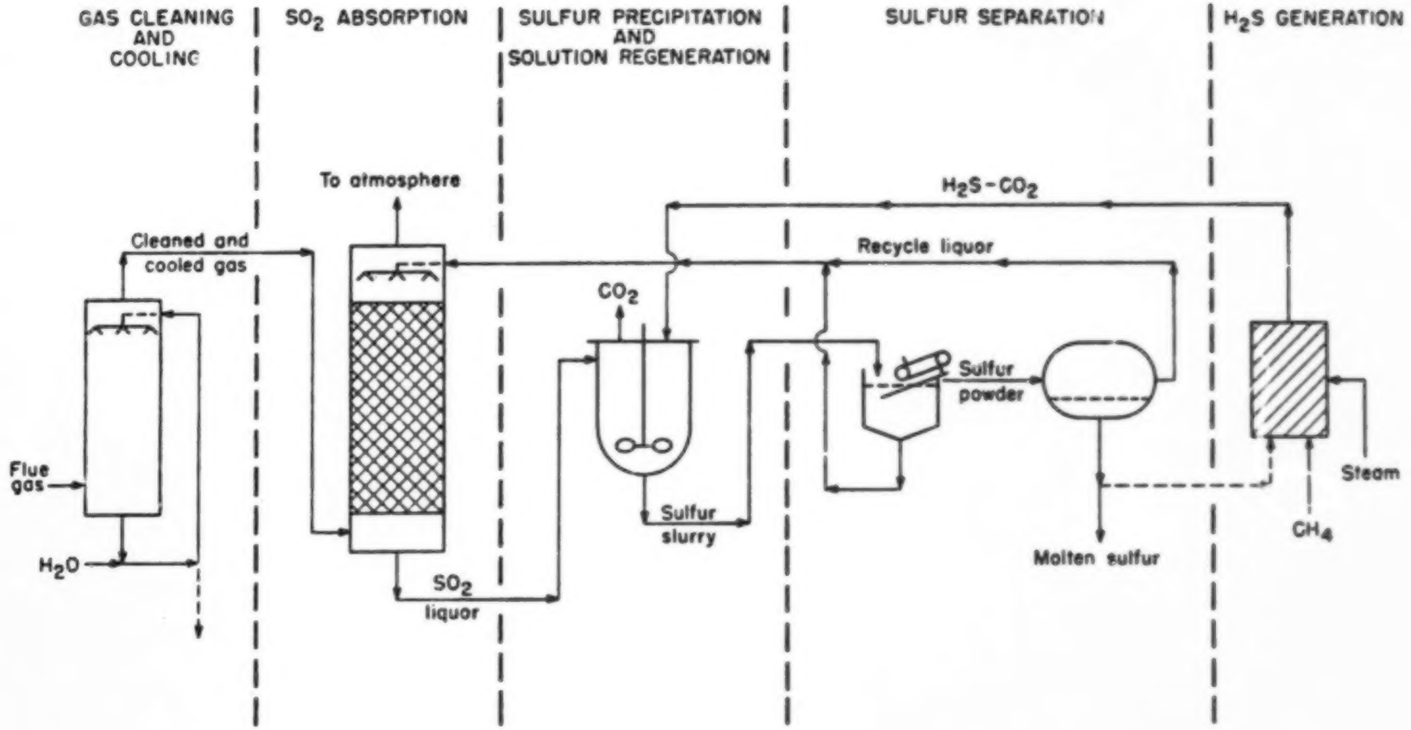

GAS CLEANING AND COOLING
SO_2 ABSORPTION
SULFUR PRECIPITATION AND SOLUTION REGENERATION
SULFUR SEPARATION
H_2S GENERATION
To atmosphere
Cleaned and cooled gas
Flue gas
H_2O
SO_2 liquor
CO_2
Sulfur slurry
H_2S-CO_2
Recycle liquor
Sulfur powder
Molten sulfur
Steam
CH_4

FIGURE B-17. GENERALIZED CITRATE PROCESS FLOWSHEET(105)

The H_2S can be generated by reacting sulfur with methane and steam:

$$CH_4 + 4\ S + 2\ H_2O \rightarrow CO_2 + 4\ H_2S.$$

The 1-3% solids slurry overflows to the effluent surge tank from where it is pumped to the conditioner tank. In the conditioner tank, the sulfur separates from the citrate slurry by floating to the surface. The regenerated citrate solution is pumped to a feed tank.

The sulfur product is withdrawn from the storage bin and pumped through a heat exchanger where the sulfur is melted at 275°F. The molten sulfur and citrate solution pass into a closed settler tank, where at 35 psi the molten sulfur settles out and is removed from the bottom. The citrate solution is returned to the absorber feed tank.

Removal Efficiency

The process is capable of up to a 99% SO_2 removal efficiency.

Raw Material and Heat Requirements

Methane, citrate, and kerosene (only for the U.S. Bureau of Mine Process) are consumed in the process. About 36,000 scf of methane are required for process heating and production of H_2S per ton of elemental sulfur recovered. About 8.4 lbs of citrate are lost per ton of sulfur recovered. And about 90 lbs of kerosene are lost per ton of sulfur recovered due to volatization from the hot sulfur slurry in the conditioner and skimmer. Various lower volatility oils can be used to minimize the evaporation loss.

By-Product

A 99.6-plus percent sulfur is generated.

Wastes

The humidifier slurry, containing fly ash and H_2SO_4, is a waste stream. It would require neutralizing with lime, thickening, and transporting to a landfill. About 0.02 lbs of sludge (50% solid) are generated per pound of SO_2 removed.

Advantages

(1) Good SO_2 removal efficiency, 90-plus percent
(2) No scaling or plugging in scrubber
(3) No major waste sludge generation
(4) Produces marketable, elemental sulfur.

Disadvantages

(1) Flue gas reheat required
(2) Requires H_2S generation gas.

Development Status

In 1968, the Bureau of Mines began investigating the citrate process at their Metallurgy Research Center in Salt Lake City. In November, 1970, a 300 cfm pilot plant went on stream at the San Manuel smelter in Arizona. It ran intermittently for 6 months. In February, 1974, the Bureau of Mines put on stream a 1000 scfm pilot plant processing off-gas from a lead smelter in Kellogg, Idaho. The Morrison-Knudsen Company, Boise, Idaho, constructed the facility. With a 5000 ppm influent SO_2 concentration, the system demonstrated up to a 99% SO_2 removal efficiency.

H_2S regeneration gas was supplied from a storage tank. In September, 1974, construction was completed on the H_2S generation plant and it is presently being tested.

On March 15, 1974, the construction of Pfizer-McKee-Peabody citrate process pilot plant was completed at Terre Haute, Indiana. The scrubber treats 2000 scfm from a coal-fired boiler with an inlet SO_2 concentration of 1000 ppm. A venturi scrubber was used in lieu of the humidifier and an impingement plate scrubbing tower replaced the packed adsorber to permit higher gas velocities. The sulfur separation was based on the flotation principle, but no hydrocarbon addition was made. Between March and September 1, the system was operated for 2330 hours. The average SO_2 removal efficiency was greater than 95%.

Capital and Operating Costs

The U.S. Bureau of Mines, Salt Lake City, Utah, estimated the capital cost of a citrate process applied to a 1000 MW coal-fired boiler burning 3 percent sulfur coal at \$36.39 x 10^6 in May, 1974.(105) The annual operating cost was estimated at \$13.87 x 10^6/yr.

Calsox Process

Developer/Manufacturer

The Monsanto Company of St. Louis, Missouri, developed the Calsox Process. They are actively engaged in marketing the process for SO_2 control to utility lockers.

Process Description

The flue gas, after passing through a forced draft fan and a humidifier/cooler, is introduced to the absorber (see Figure B-18). The absorber is of cross-flow design, although a vertical tower could be used, where the gas contacts a 0.5 weight percent enthanolamine water solution. The enthanolamine has a high affinity for SO_2 and readily absorbs it.

$$RNH_2OH + SO_2 \rightarrow RNHSOO + H_2O$$

The purified gas flows from the absorber through a mist eliminator and to a reheater before being vented to the stack. Pressure drop across the absorber is normally 2-5 inches of water.

The solution from the absorption system goes to a two-step precipitation system. First it is mixed with makeup absorbent and the liquid from the thickener. The calcium ions in the latter lead to precipitation of $CaSO_3$. The resulting stream goes to a clarifier, where the filtrate from the cake filtration is also added. The clear liquid from the clarifier is returned to the absorption system. The concentration of soluble calcium in this liquid is low enough to avoid scaling in the absorption system.

In the second stage, the clarifier bottoms are mixed with lime to complete the precipitation of $CaSO_3$. The resulting stream goes to the thickener and then to the filter. The result is a cake which contains about 50 percent water, the dry portion being composed of 85-90 percent $CaSO_3$, 10-15 percent $CaSO_4$, and about 0.5 percent ethanolamine. The $CaSO_4$,

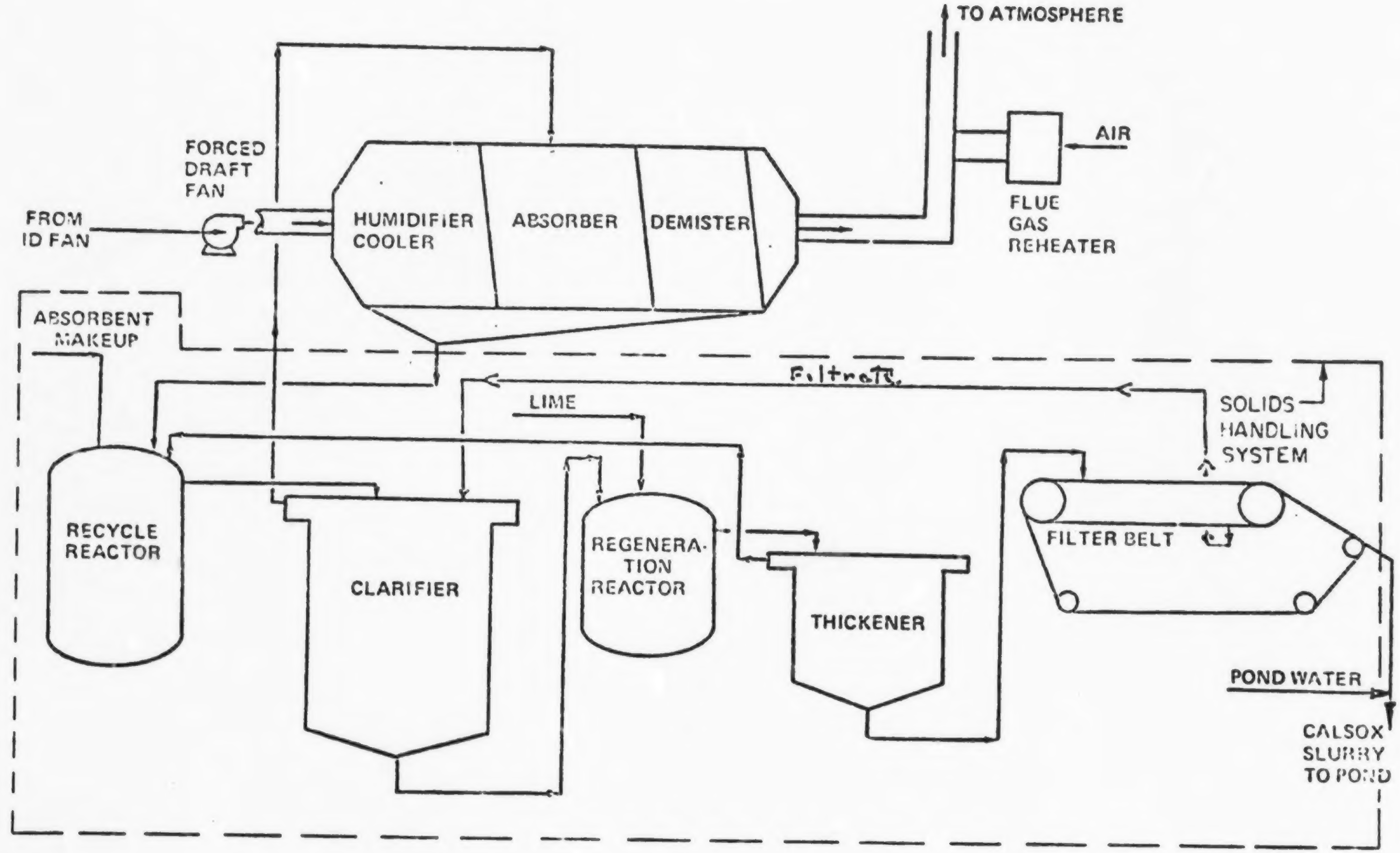

TO ATMOSPHERE
AIR
FORCED DRAFT FAN
FROM ID FAN
HUMIDIFIER COOLER
ABSORBER
DEMISTER
FLUE GAS REHEATER
ABSORBENT MAKEUP
Filtrate
LIME
SOLIDS HANDLING SYSTEM
RECYCLE REACTOR
CLARIFIER
REGENERATION REACTOR
THICKENER
FILTER BELT
POND WATER
CALSOX SLURRY TO POND

FIGURE B-18. CALSOX PROCESS(106)

an unwanted compound because of its high solubility in water, results from oxidation of the sulfite. The loss of ethanolamine in the cake is about 70 lb/hr from a 125-MW boiler. The two precipitation steps provide a countercurrent flow effect.

Removal Efficiencies

The Calsox Process can achieve a 90 percent SO_2 removal efficiency and can remove about 80 percent of the fly ash exiting the ESP. The process has essentially no NO_x removal capability.

Material and Heat Requirements

Lime is consumed in the precipitation reaction and heat is required to reheat the flue gas from the 120°F mist eliminator exit temperature to the desired 170-190°F stack gas temperature. Moreover, some ethanolamine is lost to the flue gas and filter cake.

Based on a 1:1 stoichiometry, a 125-MW facility treating flue gas with an SO_2 concentration of 3,000 ppm would consume 3.6 tons of lime (CaO) per hour. No information was available on the flue gas reheat requirements. About 2 pounds per hour of ethanolamine is lost to evaporation in the absorber; this combined with the amount lost to the filter cake yields a total loss of 72 lbs/hr for a 125-MW scrubbing facility. The cost of ethanolamine is 18¢/lb.

By-Product

No marketable by-product is generated by the Calsox Process.

Wastes

The only waste stream is the $CaSO_3$, $CaSO_4$ filter cake. A 125-MW boiler scrubbing facility would generate about 14 tons per hour of wet cake.

Advantages

(1) High SO_2 removal efficiency
(2) No plugging or scaling in scrubber.

Disadvantages

(1) Flue gas reheat necessary
(2) Waste stream generation
(3) High cost of ethanolamine.

Development Status

A 2000 scfm (3000 acfm) pilot plant was operated during the period from February through October of 1973 at a boiler owned by the Indianapolis Power and Light Company. The target of 90 percent SO_2 removal was met and several improvements in the process were worked out. A thirty-day period of uninterrupted operation was achieved in May of 1973. Prior to this (November-December, 1972) a portable, relatively unsophisticated, 5000 acfm pilot plant was operated at a utility boiler for about a month. No large-scale units have been operated to date. A design for a 125-MW boiler has been submitted to the Indianapolis Power and Light Company and is now being evaluated.

Capital and Operating Costs

For a 125 MW unit which Monsanto has designed for Indianapolis Power and Light Company, the investment, calculated to 1976, is estimated to be 9 million dollars ($72/kW).(106) This is not a minimum cost plant and includes some extra equipment such as a separator for the treated flue gas. Monsanto expects the investment to be 40-50 $/kW for large optimized plants. The operating cost was given as 3 mills/kWh for the plant.

Aqueous Carbonate Process

Developer/Manufacturer

The Atomics International (AI) Division of Rockwell International Corporation developed the Aqueous Carbonate Process in the early 1970's.

Process Description

The Aqueous Carbonate Process (ACP) employs a solution of sodium carbonate in a modified spray tower to remove SO_2 from flue gases. The spent powder is removed by a cyclone separator or an ESP and may either be disposed of (open loop) or regenerated (closed loop) to yield elemental sulfur.

The flue gas, containing SO_2 and fly ash, is first introduced into the spray tower where it co-currently contacts an atomized mist of a 4 to 20 weight percent Na_2CO_3 solution (see Figure B-19).

Instead of spray nozzles, centrifugal wheels are employed to circulate the droplets at a high velocity. The system normally operates with a liquid/gas ratio of about 1/3 gal/1000 scf. The liquid flow rate is determined by the flue gas temperature and the Na_2CO_3 concentration is determined by the SO_2 concentration in the flue gas.

The SO_2 reacts with the sodium carbonate to form sodium sulfite (Na_2SO_3), and sodium sulfate (Na_2SO_4).

$$SO_2 + NaCO_3 \rightarrow Na_2SO_3 + CO_2$$

$$Na_2SO_3 + \frac{1}{2} O_2 \rightarrow Na_2SO_4.$$

The flue gas and particulate matter leave the dryer at about 160°F, at least 20°F above the dew point, as the thermal energy of the flue gases (at about 300°F) is sufficient to vaporize the water in the spray dryer without saturating the gas. At these conditions, the flue gas requires little or no reheat.

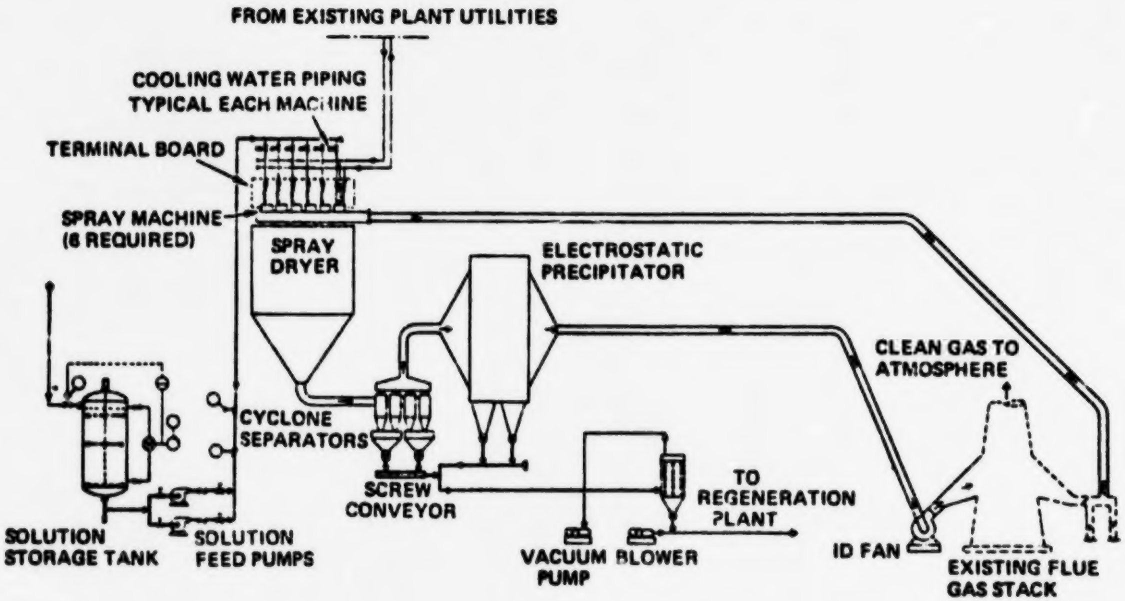

FROM EXISTING PLANT UTILITIES
COOLING WATER PIPING TYPICAL EACH MACHINE
TERMINAL BOARD
SPRAY MACHINE (6 REQUIRED)
SPRAY DRYER
ELECTROSTATIC PRECIPITATOR
CLEAN GAS TO ATMOSPHERE
CYCLONE SEPARATORS
SCREW CONVEYOR
TO REGENERATION PLANT
SOLUTION STORAGE TANK
SOLUTION FEED PUMPS
VACUUM PUMP
BLOWER
ID FAN
EXISTING FLUE GAS STACK

FIGURE B-19. SCRUBBER SUBSYSTEM LOOP(107)

The system may be operated open loop, by simply disposing of the dry waste products, or closed loop, regenerating the Na_2CO_3 reactant. Open loop operation is less costly, but the soluble Na_2SO_3 and Na_2SO_4 salts present a disposal problem. In closed loop operation, the scrubbing and regeneration systems are independent and can be uncoupled and operated separately.

The ACP regeneration system involves three chemical steps. First, the sodium sulfite and sulfate are reduced to sodium sulfide with either coke or coal at 1700°F. AI has developed a high temperature molten salt reactor for this reduction step.

$$Na_2SO_3 + 3/2\ C \rightarrow Na_2S + 3/2\ CO_2$$

$$Na_2SO_4 + 2C \rightarrow Na_2S + 2CO_2$$

Second, the sulfide is dissolved in water and treated with a CO_2-rich gas to reform the Na_2CO_3 for recycle to the scrubber and to evolve a gas rich in hydrogen sulfide. Technology similar to that used in chemical recovery processes of the pulp and paper industry is used in this step.

$$Na_2S + CO_2 + H_2O \rightarrow H_2S + 2Na^+ + CO_3^=$$

The third and final step is the conversion of the H_2S to elemental sulfur by a Claus process. The tail gas from the Claus plant is recycled to the scrubber.

In the reducer, part of the coke or coal is burned to provide the necessary heat. The molten mixture leaving the reducer goes to a quench tank, where it is dissolved to yield the "green liquor." The off gas from the reducer is used as the source of CO_2 for subsequent carbonation steps and as a source of process heat.

The green liquor is cooled and filtered to remove any carbon or flyash present. The conversion of the Na_2S in this liquor back to Na_2CO_3 involves aqueous chemical processing steps which are also used in several paper-making technologies. AI is working to develop a conversion process specifically for the ACP and considers the results obtained so far quite promising. However, as a backup for this step, they have also evaluated the Tampella process, which was developed for the pulp and paper industry in Finland and is licensed in this country by the Babcock and Wilcox Company. The Tampella process will meet all the requirements for this step of the ACP and represents proven technology and hardware. Under a contract from AI, Babcock and Wilcox recently completed an engineering study of the application of the Tampella process to this step of the ACP. The results were quite satisfactory, and B&W is now prepared to design and deliver the equipment for a full-scale modular unit.

The Tampella process for converting Na_2S into Na_2CO_3 is included in the regeneration flowsheet shown in Figure 4. The precarbonator serves to convert Na_2S into NaHS by reaction with the CO_2 in the reducer off-gas.

Precarbonator: $2Na_2S + CO_2 + H_2O \rightarrow 2NaHS + Na_2CO_3$

The carbonator produces an $NaHCO_3$ slurry by reaction of Na_2CO_3 with CO_2, also obtained from the reducer off-gas. The precarbonated liquor and the $NaHCO_3$ slurry are combined in the stripper, where steam is used to strip H_2S from the liquid phase. The stripped liquor is then pumped to a crystallizer, where $Na_2CO_3 \cdot H_2O$ is produced for recycle to the carbonator. The final product Na_2CO_3 solution is returned to the scrubber after appropriate dilution. Claus plant feed gas from this process typically contains 80-90% H_2S after condensation of the water vapor.

Carbonator: $Na_2CO_3 + H_2O + CO_2 \rightarrow 2NaHCO_3$

Stripper: $NaHS + NaHCO_3 \rightarrow H_2S + Na_2CO_3$

Crystallizer: $Na_2CO_3 + H_2O \rightarrow Na_2CO_3\ H_2O\downarrow$

Removal Efficiencies

The ACP can achieve greater than 90 percent SO_2 removal. Particulate removal is very high (> 99.8%) because of the electrostatic precipitator. The process removes some NO_2 and NO, the overall NO_x removal being about 5 percent.

Material and Heat Requirements

Atomics International maintains that the flue gas normally would not require reheating. The primary heat requirement for the closed loop system is for heating the dry carbonate, sulfite, sulfate and fly ash mixture to 1700°F in the regenerator. Coke is normally used in the regenerator, as it serves as a reducer and furnishes the necessary heat for maintaining the reaction. A 330-MW scrubbing facility would consume about 46,200 tons per year of coke for a boiler burning 3.5 wt. percent sulfur coal (107).

By-Product

The closed loop process generates elemental sulfur by-product. The 330-MW sample plant would generate about 32,000 tons of sulfur per year.(107)

Waste Streams

No liquid waste streams are produced. The open loop version of the process yields a solid product consisting of Na_2SO_3, Na_2SO_4, fly ash, and small amounts of Na_2CO_3. A 125-MW unit processing 3 percent sulfur coal and having a raw flue gas ash content of 2 grain/scf would product about 6.4 ton/hour of solid, with about 1.4 ton/hour of this being fly ash. The closed loop version of the process produces no waste streams, since the sulfur is recovered in elemental form.

Advantages

(1) No scaling or plugging
(2) Open loop version is a relatively simple process
(3) High SO_2 removal efficiency

(4) No flue gas reheat
(5) Low L/G
(6) No waste sludge disposal problem
(7) Good turndown capability
(8) Low pressure drop

Disadvantages

(1) Sodium salt disposal problem for open loop version
(2) High coal or coke consumption
(3) Closed loop version requires Claus plant

Development Status

The complete ACP has not been tested in a single installation, but both the scrubbing system and the reduction reactor have been tested separately. The rest of the regeneration system is considered proven technology.

The scrubbing portion, or open loop version, was tested in 1972. Both a 5- and 7-foot diameter spray tower were used to assess the system performance with SO_2 inlet concentrations varying from 200 to 8,000 ppm at the Mohave Power Generating Station in Laughlin, Nevada.

Reduction tests were conducted at Santa Susana on a reaction vessel equivalent to a 3-5 MW size system. The studies demonstrated good conversion efficiencies and the off-gas compositions obtained have been suitable for use in the aqueous regeneration steps.

Capital and Operating Costs

AI has made a detailed analysis of the investment and operating costs for a 330 MW unit (two 165 MW trains)(107). The total investment for the regenerative system is about 21 million dollars (63.8 $/kW) and the total operating cost is 2.8 mills/kWh. This investment includes all equipment, engineering, management, construction, startup, and shakedown costs and is based on a starting date of January, 1974. The costs include no by-product sulfur credit and no reheating cost (since reheating is considered unnecessary for most applications). The costs do include the costs of the required Claus plant. A breakdown of the utility costs is shown in Table B-22.

TABLE B-22. BREAKDOWN OF UTILITY COSTS FOR 330-MW ACP UNIT(107)

Utility	Unit Cost	mills/kWh
Electricity	10 mills/kWh	0.230
Natural Gas (or oil)	\$0.40/Mcf	0.039
Coke	\$20/ton	0.400
Cooling Water	\$0.03/Mgal	0.0087
Process Water	\$0.25/Mgal	0.0136
Boiler Feed Water	\$0.40/Mgal	0.0022
Steam	\$0.40/Mlb	0.0114
Makeup Na_2CO_3	\$50/ton	0.0076
Total Utility Cost		0.7125

APPENDIX C

ESTIMATED COSTS OF CENTRAL REGENERATION AND ACID PRODUCTION PLANT

TABLE C-1. CENTRAL REGENERATION AND ACID PLANT (1000 tons/day, 330 days/yr)

Item	Cost (mid-1973), $10^3
Capital Requirement	
Bare cost(a)	8,612
Engineering and design	--
Contractor's overhead and profit	--
Subtotal Plant Investment	8,612
Project contingency	--
Total Plant Investment	8,612
Interest during construction	388
Startup cost	1,221
Working capital	1,221
Capital Requirement	11,442
Annual Operating Cost	
Labor(b)	339
Administrative and general overhead	203
Materials and utilities(c)	2,111
Transportation of solids(d)	3,220
Additional fuel requirement	0
Local taxes and insurance	233
Gross Operating Cost	6,106
Credit	6,600
Net Operating Cost	-494
Annualized Cost	
Return on rate base	665
Federal income tax	219
Depreciation	511
Net operating cost	-494
Average Annual Cost	901
Annualized Control Cost	
$/lb S removed (e)	8.53

Footnotes to Table C-1.

(a) $\$8.2 \times 10^6$ for a 1000 ton/day plant in 1972 $(8,200)(\frac{144.1}{137.2}) = 8,612$.

This cost is the total plant investment.

(b)

Direct operation:	$123,000
Maintenance:	172,000
Supervision:	44,000
	$339,000

(c)

Power:	$ 165,000	
Boiler feed water:	21,000	
Process water:	7,000	
Cooling water:	20,000	
Fuel oil:	1,129,000	
MgO:	462,000	
Coke:	33,000	
Maintenance material:	172,000	(2 percent of TPI)
Operating supplies:	102,000	
	$2,111,000	

(d) $MgSO_3$ to be handled = 3.3×10^5 ton/yr; MgO to be handled = 1.3×10^5 ton/yr; cost of transportation (assumed) = \$7/ton; cost of transportation = \$3,220,000/yr.

(e) Sulfur removed = $(330,000)(0.98)(\frac{32}{98}) = 1.056 \times 10^5$ tons/yr.